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# SO2 Emissions in Natural Gas Production Industry Background Information for Proposed Standards

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Research Triangle Park, NC

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Research Triangle Park NC 27711

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# SO2 Emissions in Natural Gas Production Industry — Background Information for Proposed Standards

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16. ABSTRACT

Standards of performance for the control of $SO_2$ emissions from natural gas sweetening operations are being proposed under Section 111 of the Clean Air Act. This document contains background information and environmental and economic impact assessments of the regulatory alternatives considered in developing the proposed standards.

17. KEY WORDS AND DOCUMENT ANALYSIS

| a. DESCRIPTORS | b. IDENTIFIERS/OPEN ENDED TERMS | c. COSATI Field/Group |
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# $SO_2$ Emissions in Natural Gas Production Industry — Background Information for Proposed Standards

Emission Standards and Engineering Division

U.S ENVIRONMENTAL PROTECTION AGENCY
Office of Air, Noise, and Radiation
Office of Air Quality Planning and Standards
Research Triangle Park, North Carolina 27711

November 1983

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ENVIRONMENTAL PROTECTION AGENCY

Background Information
and Draft Environmental
Impact Statement for
Natural Gas Production Industry

Prepared by:

Jack R Farmer

11/9/83
(Date)

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1. The proposed standards of performance will limit emissions of sulfur dioxide from new, modified, and reconstructed onshore natural gas processing plants. Section 111 of the Clean Air Act (42 U.S.C. 7411), as amended, directs the Administrator to establish standards of performance for any category of new stationary source of air pollution that ". . . causes or contributes significantly to air pollution which may reasonably be anticipated to endanger public health or welfare." The proposed standards of performance are expected to affect mostly the states where smackover, permian or overthrust belt regional gas fields are located.

2. Copies of this document have been sent to the following Federal Departments: Office of Management and Budget, Labor, Health and Human Services, Defense, Transportation, Agriculture, Commerce, Interior, and Energy; the National Science Foundation; the Council on Environmental Quality; members of the State and Territorial Air Pollution Program Administrators; the Association of Local Air Pollution Control Officials; EPA Regional Administrators; and other interested parties.

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# 1. SUMMARY

Standards of performance for new stationary sources are established under Section 111 of the Clean Air Act (42 U.S.C. 7411), as amended in 1977. Section 111 directs the Administrator to establish standards of performance for any category of new stationary source of air pollution which ". . . causes or contributes significantly to air pollution which may reasonably be anticipated to endanger public health or welfare." This background information document (BID) supports the proposed standards, which would control sulfur dioxide ($SO_2$) emissions from onshore natural gas processing facilities.

The natural gas production industry is involved in producing pipeline quality (sweet) natural gas. Over 75 percent of the onshore production, and almost all of the offshore production is sweet. The remaining fraction, however, contains acid gases and is routinely sweetened to reduce the hydrogen sulfide ($H_2S$) and carbon dioxide ($CO_2$) content to levels that are acceptable for pipeline distribution. Elemental sulfur may be recovered from $H_2S$ in the separated acid gas ($H_2S$ and $CO_2$) stream for those onshore processing facilities that sweeten the sour natural gas. The residual $H_2S$ is oxidized to $SO_2$, and released to the atmosphere.

## 1.1 REGULATORY ALTERNATIVES

To evaluate the environmental, economic, and energy impacts associated with implementation of a standard for the natural gas production industry, the Administrator has examined a number of regulatory alternatives for controlling $SO_2$ emissions. The six regulatory alternatives selected for evaluation are summarized as follows:

- Regulatory Alternative I – No standards of performance would be promulgated for onshore natural gas processing facilities. This alternative assumes that current state and local regulations are applied to the industry (baseline control). Baseline control technologies range from incineration (zero percent $SO_2$ reduction) to the Claus process (2- or 3-stage, 93.01 to 97.31 percent start-of-run $SO_2$ reduction) depending on plant size (sulfur feed rate) and acid gas $H_2S/CO_2$ ratio.

- Regulatory Alternatives II through VI – Regulatory Alternatives II through VI apply five control technologies in different combinations to differing plant sizes resulting in progressively more stringent levels of $SO_2$ emissions reduction. The technologies include the Claus process (2-stage, 93.01 to 96.31 percent start-of-run $SO_2$ reduction; 3-stage, 94.72 to 97.31 percent start-of-run $SO_2$ reduction); the Recycle Selectox process (2-stage, 80.68 to 92.36 percent start-of-run $SO_2$ reduction; 3-stage, 83.61 to 95.12 percent start-of-run $SO_2$ reduction); the Sulfreen process (97.94 to 98.82 percent start-of-run $SO_2$ reduction); The Shell Claus Offgas Treatment process (99.89 to 99.99 percent start-of-run $SO_2$ reduction) and the Beavon Sulfur Recovery process (99.89 to 99.99 percent start-of-run $SO_2$ reduction). Regulatory Alternatives II through VI progressively increase in cost per megagram of $SO_2$ reduced. The impacts of each regulatory alternative are assessed through analyses of their impacts on individual plant sizes. Nationwide impacts are determined based on a projected distribution of plant sizes to be constructed.

## 1.2 ENVIRONMENTAL IMPACT

The environmental and energy impacts of the regulatory alternatives are summarized in Table 1-1. Regulatory Alternative I has the only adverse air impact while Alternatives III through VI produce significant emission reductions. No adverse water or solid waste impacts and only negligible adverse energy impacts are associated with any of the regulatory alternatives. The environmental and energy impacts are discussed in detail in Chapter 7 of the BID.

Table 1-1. ASSESSMENT OF ENVIRONMENTAL, ENERGY, AND ECONOMIC IMPACTS FOR EACH REGULATORY ALTERNATIVE CONSIDERED FOR THE NATURAL GAS PRODUCTION INDUSTRY[a]

| Administrative alternative | Air impact | Water impact | Solid waste impact | Energy impact | Economic impact |
|---|---|---|---|---|---|
| Regulatory Alternative I | -3** | 0 | 0 | 0 | 0 |
| Regulatory Alternative II | +2** | 0 | 0 | -1** | 0 |
| Regulatory Alternative III | +3** | 0 | 0 | -1** | -1** |
| Regulatory Alternative IV | +4** | 0 | 0 | -1** | -2** |
| Regulatory Alternative V | +4** | 0 | 0 | -1** | -3** |
| Regulatory Alternative VI | +4** | 0 | 0 | -1** | -5** |

[a] KEY:
+ Beneficial impact
- Adverse impact
0 No impact
1 Negligible impact
2 Small impact
3 Moderate impact
4 Large impact
5 Very large impact
* Short term impact
** Long term impact
*** Irreversible impact

## 1.3 ECONOMIC IMPACT

The estimated economic impacts also are summarized in Table 1-1. Regulatory Alternatives V and VI have large adverse economic impacts. The economic impacts of Regulatory Alternatives I through IV range from no impact to a small adverse impact. Regulatory Alternative IV has the greatest beneficial air quality impact without a large adverse economic impact. The economic impacts are discussed in detail in Chapter 9 and cost analyses are discussed in Chapter 8 of the BID.

# 2. INTRODUCTION

## 2.1 BACKGROUND AND AUTHORITY FOR STANDARDS

Before standards of performance are proposed as a Federal regulation, air pollution control methods available to the affected industry and the associated costs of installing and maintaining the control equipment are examined in detail. Various levels of control based on different technologies and degrees of efficiency are expressed as regulatory alternatives. Each of these alternatives is studied by EPA as a prospective basis for a standard. The alternatives are investigated in terms of their impacts on the economics and well-being of the industry, the impacts on the national economy, and the impacts on the environment. This document summarizes the information obtained through these studies so that interested persons will be privy to the information considered by EPA in the development of the proposed standard.

Standards of performance for new stationary sources are established under Section 111 of the Clean Air Act (42 U.S.C. 7411) as amended, hereinafter referred to as the Act. Section 111 directs the Administrator to establish standards of performance for any category of new stationary source of air pollution which ". . . causes, or contributes significantly to air pollution which may reasonably be anticipated to endanger public health or welfare."

The Act requires that standards of performance for stationary sources reflect, ". . . the degree of emission reduction achievable which (taking into consideration the cost of achieving such emission reduction, and any nonair quality health and environmental impact and energy requirements) the Administrator determines has been adequately demonstrated for that category of sources." The standards apply only to stationary sources, the construction or modification of which commences after regulations are proposed by publication in the Federal Register.

The 1977 amendments to the Act altered or added numerous provisions that apply to the process of establishing standards of performance.

1. EPA is required to review the standards of performance every 4 years and, if appropriate, revise them.

2. EPA is authorized to promulgate a standard based on design, equipment, work practice, or operational procedures when a standard based on emission levels is not feasible.

3. The term "standards of performance" is redefined, and a new term "technological system of continuous emission reduction" is defined. The new definitions clarify that the control system must be continuous and may include a low- or non-polluting process or operation.

4. The time between the proposal and promulgation of a standard under section 111 of the Act may be extended to 6 months.

Standards of performance, by themselves, do not guarantee protection of health or welfare because they are not designed to achieve any specific air quality levels. Rather, they are designed to reflect the degree of emission limitation achievable through application of the best adequately demonstrated technological system of continuous emission reduction, taking into consideration the cost of achieving such emission reduction, any nonair-quality health and environmental impacts, and energy requirements.

Congress had several reasons for including these requirements. First, standards with a degree of uniformity are needed to avoid situations where some States may attract industries by relaxing standards relative to other States. Second, stringent standards enhance the potential for long-term growth. Third, stringent standards may help achieve long-term cost savings by avoiding the need for more expensive retrofitting when pollution ceilings may be reduced in the future. Fourth, certain types of standards for coal-burning sources can adversely affect the coal market by driving up the price of low-sulfur coal or effectively excluding certain coals from the reserve base because their untreated pollution potentials are high. Congress does not intend that new source performance standards contribute to these problems. Fifth, the standard-setting process should create incentives for improved technology.

Promulgation of standards of performance does not prevent State or local agencies from adopting more stringent emission limitations for the same sources. States are free under Section 116 of the Act to establish even more stringent emission limits than those established under Section 111 or those necessary to attain or maintain the National Ambient Air Quality Standards (NAAQS) under Section 110. Thus, new sources may in some cases be subject to limitations more stringent than standards of performance under Section 111, and prospective owners and operators of new sources should be aware of this possibility in planning for such facilities.

A similar situation may arise when a major emitting facility is to be constructed in a geographic area that falls under the prevention of significant deterioration of air quality provisions of Part C of the Act. These provisions require, among other things, that major emitting facilities to be constructed in such areas are to be subject to best available control technology. The term Best Available Control Technology (BACT), as defined in the Act, means

> . . . an emission limitation based on the maximum degree of reduction of each pollutant subject to regulation under this Act emitted from, or which results from, any major emitting facility, which the permitting authority, on a case-by-case basis, taking into account energy, environmental, and economic impacts and other costs, determines is achievable for such facility through application of production processes and available methods, systems, and techniques, including fuel cleaning or treatment or innovative fuel combustion techniques for control of each such pollutant. In no event shall application of 'best available control technology' result in emissions of any pollutants which will exceed the emissions allowed by any applicable standard established pursuant to Sections 111 or 112 of this Act. (Section 169(3))

Although standards of performance are normally structured in terms of numerical emission limits where feasible, alternative approaches are sometimes necessary. In some cases physical measurement of emissions from a new source may be impractical or exorbitantly expensive. Section 111(h) provides that the Administrator may promulgate a design or equipment standard in those cases where it is not feasible to prescribe or enforce a standard of performance. For example, emissions of hydrocarbons from storage vessels for petroleum liquids are greatest during tank filling. The nature of the emissions, high concentrations for short periods during filling and low concentrations for longer periods during storage, and the configuration of storage tanks make direct emission measurement impractical. Therefore, a more practical approach to standards of performance for storage vessels has been equipment specification.

In addition, Section 111(j) authorizes the Administrator to grant waivers of compliance to permit a source to use innovative continuous emission control technology. In order to grant the waiver, the Administrator must find: (1) a substantial likelihood that the technology will produce greater emission reductions than the standards require or an equivalent reduction at lower economic energy or environmental cost; (2) the proposed system has not been adequately demonstrated; (3) the technology will not cause or contribute to an unreasonable risk to the public health, welfare, or safety; (4) the governor of the State where the source is located consents; and (5) the waiver will not prevent the attainment or maintenance of any ambient standard. A waiver may have conditions attached to assure the source will not prevent attainment of any NAAQS. Any such condition will have the force of a performance standard. Finally, waivers have definite end dates and may be terminated earlier if the conditions are not met or if the system fails to perform as expected. In such a case, the source may be given up to 3 years to meet the standards with a mandatory progress schedule.

## 2.2 SELECTION OF CATEGORIES OF STATIONARY SOURCES

Section 111 of the Act directs the Administrator to list categories of stationary sources. The Administrator ". . . shall include a category of sources in such list if in his judgement it causes, or contributes significantly to, air pollution which may reasonably be anticipated to endanger public health or welfare." Proposal and promulgation of standards of performance are to follow.

Since passage of the Clean Air Amendments of 1970, considerable attention has been given to the development of a system for assigning priorities to various source categories. The approach specifies areas of interest by considering the broad strategy of the Agency for implementing the Clean Air Act. Often, these "areas" are actually pollutants emitted by stationary sources. Source categories that emit these pollutants are evaluated and ranked by a process involving such factors as: (1) the level of emission control (if any) already required by State regulations, (2) estimated levels of control that might be required from standards of performance for the source category, (3) projections of growth and replacement of existing facilities for the source category, and (4) the estimated incremental amount of air pollution that could be prevented in a preselected future year by standards of performance for the source category. Sources for which new source performance standards were promulgated or under development during 1977, or earlier, were selected on these criteria.

The Act amendments of August 1977 establish specific criteria to be used in determining priorities for all major source categories not yet listed by EPA. These are: (1) the quantity of air pollutant emissions that each such category will emit, or will be designed to emit; (2) the extent to which each such pollutant may reasonably be anticipated to endanger public health or welfare; and (3) the mobility and competitive nature of each such category of sources and the consequent need for nationally applicable new source standards of performance.

The Administrator is to promulgate standards for these categories according to the schedule referred to earlier.

In some cases it may not be feasible immediately to develop a standard for a source category with a high priority. This might happen when a program of research is needed to develop control techniques or because techniques for sampling and measuring emissions may require refinement. In the developing of standards, differences in the time required to complete the necessary investigation for different source categories must also be considered. For example, substantially more time may be necessary if numerous pollutants must be investigated from a single source category. Further, even late in the development process the schedule for completion of a standard may change. For example, inablility to obtain emission data from well-controlled sources in time to pursue the development process in a systematic fashion may force a change in scheduling. Nevertheless, priority ranking is, and will continue to be, used to establish the order in which projects are initiated and resources assigned.

After the source category has been chosen, the types of facilities within the source category to which the standard will apply must be determined. A source category may have several facilities that cause air pollution, and emissions from some of these facilities may vary from insignificant to very expensive to control. Economic studies of the source category and of applicable control technology may show that air pollution control is better served by applying standards to the more severe pollution sources. For this reason, and because there is no adequately demonstrated system for controlling emissions from certain facilities, standards often do not apply to all facilities at a source. For the same reasons, the standards may not apply to all air pollutants emitted. Thus, although a source category may be selected to be covered by a standard of performance, not all pollutants or facilities within that source category may be covered by the standards.

## 2.3 PROCEDURE FOR DEVELOPMENT OF STANDARDS OF PERFORMANCE

Standards of performance must (1) realistically reflect best demonstrated control practice; (2) adequately consider the cost, the nonair-quality health and environmental impacts, and the energy requirements of such control; (3) be applicable to existing sources that are modified or reconstructed as well as new installations; and (4) meet these conditions for all variations of operating conditions being considered anywhere in the country.

The objective of a program for developing standards is to identify the best technological system of continuous emission reduction that has been adequately demonstrated. The standard-setting process involves three principal phases of activity: (1) information gathering, (2) analysis of the information, and (3) development of the standard of performance.

During the information-gathering phase, industries are queried through a telephone survey, letters of inquiry, and plant visits by EPA representatives. Information is also gathered from many other sources to provide reliable data that characterize the pollutant emissions from well-controlled existing facilities.

In the second phase of a project, the information about the industry and the pollutants emitted is used in analytical studies. Hypothetical "model plants" are defined to provide a common basis for analysis. The model plant definitions, national pollutant emission data, and existing State regulations governing emissions from the source category are then used in establishing "regulatory alternatives." These regulatory alternatives are essentially different levels of emission control.

EPA conducts studies to determine the impact of each regulatory alternative on the economics of the industry and on the national economy, on the environment, and on energy consumption. From several possibly applicable alternatives, EPA selects the single most plausible regulatory alternative as the basis for a standard of performance for the source category under study.

In the third phase of a project, the selected regulatory alternative is translated into a standard of performance, which, in turn, is written in the form of a Federal regulation. The Federal regulation, when applied to newly constructed plants, will limit emissions to the levels indicated in the selected regulatory alternative.

As early as is practical in each standard-setting project, EPA representatives discuss the possibilities of a standard and the form it might take with members of the National Air Pollution Control Techniques Advisory Committee. Industry representatives and other interested parties also participate in these meetings.

The information acquired in the project is summarized in the Background Information Document (BID). The BID, the standard, and a preamble explaining the standard are widely circulated to the industry being considered for control, environmental groups, other government agencies, and offices within EPA. Through this extensive review process, the points of view of expert reviewers are taken into consideration as changes are made to the documentation.

A "proposal package" is assembled and sent through the offices of EPA Assistant Administrators for concurrence before the proposed standard is officially endorsed by the EPA Administrator. After being approved by the EPA Administrator, the preamble and the proposed regulation are published in the Federal Register.

As a part of the Federal Register announcement of the proposed regulation, the public is invited to participate in the standard-setting process. EPA invites written comments on the proposal and also holds a public hearing to discuss the proposed standard with interested parties. All public comments are summarized and incorporated into a second volume of the BID. All information reviewed and generated in studies in support of the standard of performance is available to the public in a "docket" on file in Washington, D. C.

Comments from the public are evaluated, and the standard of performance may be altered in response to the comments.

The significant comments and EPA's position on the issues raised are included in the "preamble" of a "promulgation package," which also contains the draft of the final regulation. The regulation is then subjected to another round of review and refinement until it is approved by the EPA Administrator. After the Administrator signs the regulation, it is published as a "final rule" in the Federal Register.

## 2.4 CONSIDERATION OF COSTS

Section 317 of the Act requires an economic impact assessment with respect to any standard of performance established under Section 111 of the Act. The assessment is required to contain an analysis of: (1) the costs of compliance with the regulation, including the extent to which the cost of compliance varies depending on the effective date of the regulation and the development of less expensive or more efficient methods of compliance; (2) the potential inflationary or recessionary effects of the regulation; (3) the effects the regulation might have on small business with respect to competition; (4) the effects of the regulation on consumer costs; and (5) the effects of the regulation on energy use. Section 317 also requires that the economic impact assessment be as extensive as practicable.

The economic impact of a proposed standard upon an industry is usually addressed both in absolute terms and in terms of the control costs that would be incurred as a result of compliance with typical, existing State control regulations. An incremental approach is necessary because both new and existing plants would be required to comply with

State regulations in the absence of a Federal standard of performance. This approach requires a detailed analysis of the economic impact from the cost differential that would exist between a proposed standard of performance and the typical State standard.

Air pollutant emissions may cause water pollution problems, and captured potential air pollutants may pose a solid waste disposal problem. The total environmental impact of an emission source must, therefore, be analyzed and the costs determined whenever possible.

A thorough study of the profitability and price-setting mechanisms of the industry is essential to the analysis so that an accurate estimate of potential adverse economic impacts can be made for proposed standards. It is also essential to know the capital requirements for pollution control systems already placed on plants so that the additional capital requirements necessitated by these Federal standards can be placed in proper perspective. Finally, it is necessary to assess the availability of capital to provide the additional control equipment needed to meet the standards of performance.

## 2.5 CONSIDERATION OF ENVIRONMENTAL IMPACTS

Section 102(2)(C) of the National Environmental Policy Act (NEPA) of 1969 requires Federal agencies to prepare detailed environmental impact statements on proposals for legislation and other major Federal actions significantly affecting the quality of the human environment. The objective of NEPA is to build into the decisionmaking process of Federal agencies a careful consideration of all environmental aspects of proposed actions.

In a number of legal challenges to standards of performance for various industries, the United States Court of Appeals for the District of Columbia Circuit has held that environmental impact statements need not be prepared by the Agency for proposed actions under Section 111 of the Clean Air Act. Essentially, the Court of Appeals has determined that the best system of emission reduction requires the Administrator to take into account counter-productive environmental effects of a proposed standard, as well as economic costs to the industry. On this basis, therefore, the Court established a narrow exemption from NEPA for EPA determination under Section 111.

In addition to these judicial determinations, the Energy Supply and Environmental Coordination Act (ESECA) of 1974 (PL-93-319) specifically exempted proposed actions under the Clean Air Act from NEPA requirements. According to Section 7(c)(1), "No action taken under the Clean Air Act shall be deemed a major Federal action significantly affecting the quality of the human environment within the meaning of the National Environmental Policy Act of 1969" (15 U.S.C. 793(c)(1)).

Nevertheless, the Agency has concluded that the preparation of environmental impact statements could have beneficial effects on certain regulatory actions. Consequently, although not legally required to do so by Section 102(2)(C) of NEPA, EPA has adopted a policy requiring that environmental impact statements be prepared for various regulatory actions, including standards of performance developed under Section 111 of the Act. This voluntary preparation of environmental impact statements, however, in no way legally subjects the Agency to NEPA requirements.

To implement this policy, a separate section in this document is devoted solely to an analysis of the potential environmental impacts associated with the proposed standards. Both adverse and beneficial impacts in such areas as air and water pollution, increased solid waste disposal, and increased energy consumption are discussed.

## 2.6 IMPACT ON EXISTING SOURCES

Section 111 of the Act defines a new source as ". . . any stationary source, the construction or modification of which is commenced . . ." after the proposed standards are published. An existing source is redefined as a new source if "modified" or "reconstructed" as defined in amendments to the general provisions of Subpart A of 40 CFR Part 60, which were promulgated in the Federal Register on December 16, 1975 (40 FR 58416).

Promulgation of a standard of performance requires States to establish standards of performance for existing sources in the same industry under Section 111 (d) of the Act if the standard for new sources limits emissions of a designated pollutant (i.e., a pollutant for which air quality criteria have not been issued under Section 108 or which has not been listed as a hazardous pollutant under Section 112). If a State does not act, EPA must establish such standards. General provisions outlining procedures for control of existing sources under Section 111(d) were promulgated on November 17, 1975, as Subpart B of 40 CFR Part 60 (40 FR 53340).

## 2.7 REVISION OF STANDARDS OF PERFORMANCE

Congress was aware that the level of air pollution control achievable by any industry may improve with technological advances. Accordingly, section 111 of the Act provides that the Administrator ". . . shall, at least every 4 years, review and, if appropriate, revise . . ." the standards. Revisions are made to assure that the standards continue to reflect the best systems that become available in the future. Such revisions will not be retroactive, but will apply to stationary sources constructed or modified after the proposal of the revised standards.

# 3. THE NATURAL GAS PRODUCTION INDUSTRY

The crude oil and natural gas production industry involves a large number of similar yet distinct industrial processes linked together in widely differing combinations to meet a common purpose. That purpose is to remove hydrocarbons and associated compounds from subterranean deposits of oil and gas and to produce marketable products for industrial, commercial, and residential use. On a detailed level, the methods of removing the oil and gas from the earth, the process operations employed, and the products produced all vary as greatly as the types of geologic formations containing oil and gas, the properties of the well materials obtained, and the markets for which the products are produced. However, on a more general level, the industry can be described by several operations and installations distinguished from one another primarily by the purpose or function they serve.

## 3.1 GENERAL DESCRIPTION

Three basic operations within the crude oil and natural gas production industry that are important to the control of sulfur compounds emissions into the atmosphere are production operations, sweetening operations, sulfur recovery operations, and incineration operations. Figure 3-1 shows a generalized flow diagram of major industry operations and the emission sources associated with each operation.

Production operations for the crude oil and natural gas industry consist of bringing reservoir fluids to the surface and providing the field processing or treatment required to produce commercial products. The purpose of the first major operation, well drilling, is to produce a point of access to the subterranean deposit for exploration, oil and gas production, water injection, gas injection, or other purposes.

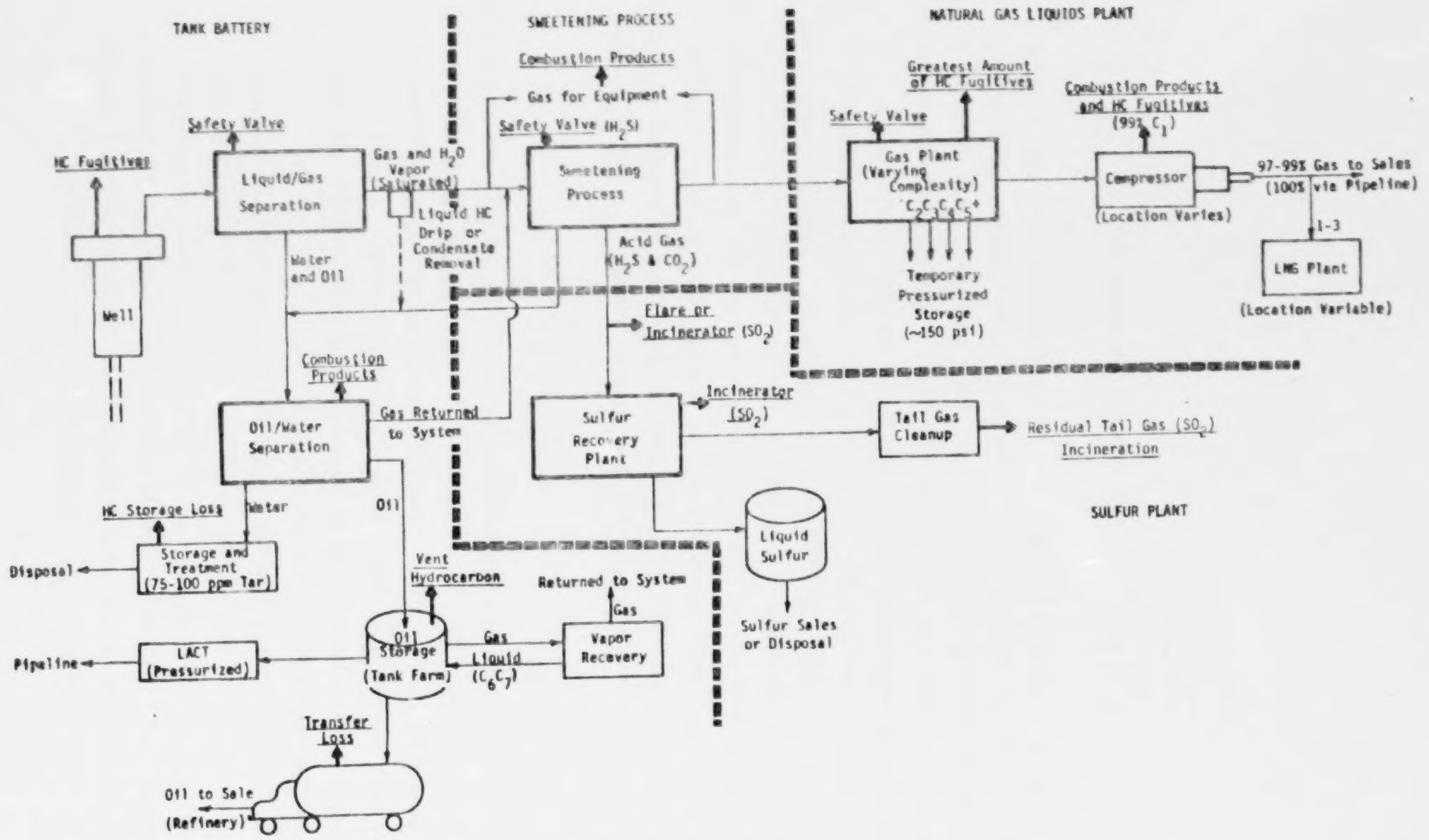
TANK BATTERY
SWEETENING PROCESS
NATURAL GAS LIQUIDS PLANT
SULFUR PLANT
HC Fugitives
Well
Safety Valve
Liquid/Gas Separation
Gas and $H_2O$ Vapor (Saturated)
Liquid HC Drip or Condensate Removal
Water and Oil
Combustion Products
Oil/Water Separation
Gas Returned to System
Oil
Water
HC Storage Loss
Storage and Treatment (75-100 ppm Tar)
Disposal
Vent Hydrocarbon
Oil Storage (Tank Farm)
Gas
Liquid ($C_6C_7$)
Vapor Recovery
Returned to System
LACT (Pressurized)
Pipeline
Transfer Loss
Oil to Sale (Refinery)
Combustion Products
Gas for Equipment
Safety Valve ($H_2S$)
Sweetening Process
Acid Gas ($H_2S$ & $CO_2$)
Flare or Incinerator ($SO_2$)
Sulfur Recovery Plant
Incinerator ($SO_2$)
Liquid Sulfur
Sulfur Sales or Disposal
Tail Gas Cleanup
Residual Tail Gas ($SO_2$) Incineration
Safety Valve
Greatest Amount of HC Fugitives
Gas Plant (Varying Complexity)
$C_2C_3C_4C_5$+
Temporary Pressurized Storage (~150 psi)
Combustion Products and HC Fugitives (99% $C_1$)
Compressor
(Location Varies)
97-99% Gas to Sales (100% via Pipeline)
1-3
LNG Plant
(Location Variable)

Figure 3-1. Production and processing operations and associated emissions in the natural gas production industry.

The next major processing operations applied to natural gas after it has been removed from gas wells or gas/oil wells are the gas-liquid separation operation and the sweetening operation. All the onshore natural gas that is "sour" is sweetened. Sweetening is the removal of hydrogen sulfide ($H_2S$) and carbon dioxide ($CO_2$) gases present in "sour" natural gas. Sour gas is natural gas with a $H_2S$ concentration greater than 0.25 grains per 100 standard cubic feet.[1] Sour natural gas contains widely differing concentrations of $H_2S$ and $CO_2$ and trace amounts of organic sulfur compounds such as mercaptans (RSH). $H_2S$ is rarely less than 95 percent of the total sulfur content.[2]

Removal of $H_2S$ and $CO_2$ from natural gas streams is necessary to make the natural gas suitable for consumer use. Specifications on the degree of $H_2S$ removal allowed are essentially standard and uniform across the United States. The maximum $H_2S$ content is 0.25 grains of $H_2S$ per 100 standard cubic feet (4 ppmv $H_2S$) of sweetened natural gas. This standard remains unchanged as the criterion for "sweet gas" in most natural gas applications. Sweet natural gas is also termed "residue gas" in the industry. Generally, $CO_2$ may be transported in natural gas streams as long as the quantity of $CO_2$ does not reach the point of seriously lowering the heating value of the gas.

Currently, there are a number of processes used to sweeten sour natural gases. These processes are listed and described briefly in Table 3-1. Several processes are employed frequently for selective absorption of $H_2S$. A simplified flow diagram for a typical gas sweetening facility is shown in Figure 3-2. Amine treating of sour natural gas for the removal of $H_2S$ and $CO_2$ is probably the most widely utilized process for sweetening the sour gas in the industry. This process involves scrubbing the gas with amine solutions that absorb $H_2S$ and $CO_2$. Regeneration (stripping operation) of this absorbing solution produces an acid gas stream, containing $H_2S$, $CO_2$, the saturated amount of water vapor, and negligible amounts of hydrocarbons. This acid gas stream is either flared, incinerated, or processed further in a sulfur recovery facility to recover liquid elemental sulfur from $H_2S$ in the acid gas stream.

## Table 3-1. MAJOR SWEETENING PROCESSES

| Process | Status of commercialization[a] | Process category | Comments[b] |
|---|---|---|---|
| 1. Ethanolamine processes | | | |
| a. Monoethanolamine (MEA) | Well established technology (being) displaced) | Chemical absorption | Being replaced by more efficient systems, particularly for the treatment of high pressure natural gas. Still preferred solvent for gas streams containing low concentrations of acid gas ($H_2S$ and $CO_2$) and essentially no minor contaminants such as COS and $CS_2$.[3] |
| b. Methyldiethanolamine (MDEA) | Established technology | Chemical absorption | Generally used for selective absorption of $H_2S$ in the presence of $CO_2$. |
| c. Diethanolamine (DEA) | Well established technology | Chemical absorption | Utilized when acid gas ($H_2S$ and $CO_2$) is less than 25-30%. Split flow DEA more commonly used when acid gas concentration is above 30%. |
| d. Diglycolamine (DGA) | Established technology | Chemical absorption | |
| e. Diisopropanolamine (DIPA) | Established technology | Chemical absorption | Selective for $H_2S$ and in the presence of $CO_2$. |
| f. Triethanolamine (TEA) | | Chemical absorption | Technology was not commercially adopted. |
| 2. Sulfinol | Established technology | Chemical & physical absorption | Best application is in a gas stream having high total acid gas partial pressures and with relatively high ratios of $H_2S_1$ to $CO_2$ (i.e., ratio 1:1 or higher). Elevated pressure required for efficient removal of $H_2S$. |
| 3. Selexol | Established technology | Physical absorption | Lower initial plant costs and operating costs than MEA or $K_2CO_3$.[1] Elevated pressure required for efficient removal of $H_2S$. |

(continued)

Table 3-1. Concluded

| Process | Status of commercialization[a] | Process category | Comments[b] |
|---|---|---|---|
| 4. Hot potassium carbonate ($K_2CO_3$) | Well established technology (Benfield process) | Chemical absorption | Very effective where 8 to 50% acid gases are present in large quantities. Unsuitable for gas mixtures containing little or no $CO_2$. |
| 5. Stretford process | Established technology | Catalytic chemical reaction | This process yields elemental sulfur and is a combination sweetening process/sulfur recovery plant. The process does not remove organic sulfur compounds. |
| 6. Flour solvent process | Established technology | Physical absorption | More economical than activated hot potassium carbonate.[3] Elevated pressure required for efficient removal of $H_2S$. |
| 7. Purisol | Established technology | Physical absorption | Particularly suitable for selective $H_2S$ absorption in the presence of $CO_2$.[3] Elevated pressure required for efficient removal of $H_2S$. |
| 8. Iron oxide (sponge) process | Technology is being displaced | Catalytic chemical reaction (regenerative) | Recently developed wet purification processes are replacing oxide box purifiers. The liquid processes have substantial economic advantages over the iron oxide processes.[3] Disposal of the spent iron oxide can be a problem. However, in some situations, it is the most viable process. |
| 9. Nickel chloride process | | | Texaco's nickel chloride sweetening process. |

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[a] In general, if more than five facilities exist, the technology is considered to be established.

[b] The selection of a sweetening process depends upon $H_2S$ and $CO_2$ contents and also the type and quantity of heavier hydrocarbons.

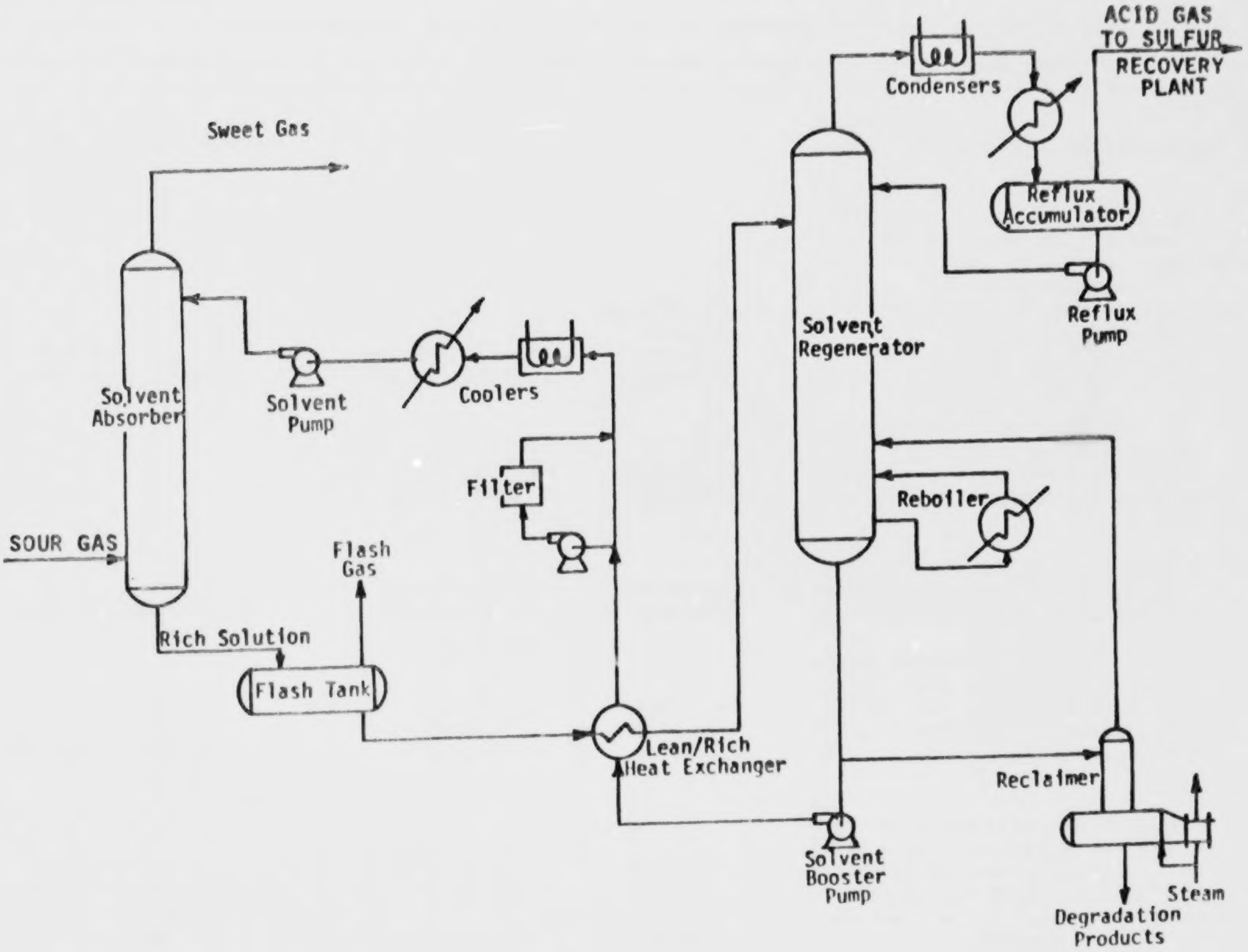

Sweet Gas
ACID GAS
TO SULFUR
RECOVERY
PLANT
Condensers
Reflux
Accumulator
Reflux
Pump
Solvent
Absorber
Solvent
Pump
Coolers
Solvent
Regenerator
Filter
Reboiler
SOUR GAS
Flash
Gas
Rich Solution
Flash Tank
Lean/Rich
Heat Exchanger
Reclaimer
Solvent
Booster
Pump
Steam
Degradation
Products

Figure 3-2. Typical sour natural gas amine sweetening facility.

## 3.2 SULFUR RECOVERY OPERATIONS AND SULFUR DIOXIDE ($SO_2$) EMISSIONS

The Claus process is the most widely used process for converting $H_2S$ in acid gases separated from natural gas into elemental sulfur. The feed to the Claus plant is an acid gas stream removed from the original sour natural gas stream by a sweetening process. Sulfur recovery efficiencies are greater (i.e., $SO_2$ emissions are lower) with higher concentrations of $H_2S$ in the feed stream. A number of Claus sulfur recovery process configurations are used by the industry to accommodate various concentrations of $H_2S$ in the acid gas feed. Currently, over 250 Claus sulfur recovery plants are operating worldwide in petroleum refineries and onshore natural gas processing facilities. Plant capacities range from less than 5 to over 1,000 metric tons of sulfur feed per day.

The Stretford process is a wet chemistry process for converting $H_2S$ to elemental sulfur. Originally the process was used for removing $H_2S$ from coal-derived gases; subsequently it has been applied to gases from various sources including refinery gases and natural gas streams. The Stretford process is highly effective in removing low concentrations of $H_2S$ from contaminated gas streams. This process is commercially available and is demonstrated in more than 50 plants throughout the world. However, the Stretford process has been utilized on a limited scale in refinery and onshore natural gas processing operations in the United States.

### 3.2.1 Claus Sulfur Recovery Process

Claus sulfur recovery plants convert $H_2S$ to elemental sulfur by gas phase reactions at 8 to 10 psig initial pressure (with tail gas systems as much as 20 psig) and at high temperatures 190°C to 330°C (375°F to 625°F), using catalysts in stages. Conversion of the $H_2S$ in the acid gas feed to elemental sulfur can be from 70 to 97 percent complete, depending on the concentration of $H_2S$ in the acid gas feed stream, the type of process configuration, and the number of reaction stages employed. Several side reactions occur during the Claus process, some of which can produce such compounds as carbonyl sulfide (COS) and carbon disulfide ($CS_2$). Unconverted sulfur compounds that appear in the tail gases from Claus plants are thermally oxidized and emitted to the atmosphere as $SO_2$ or processed further for sulfur recovery in a tail gas cleanup unit.

There are four main variations of the Claus process, differing primarily in the way in which the heat balance is maintained:

1. the straight-through process,
2. the split-flow process,
3. the split-flow process with preheating of feed stream, and
4. the sulfur-recycle process.

The straight-through process is utilized when the $H_2S$ concentration in the acid gas feed is high (greater than 50 mole percent). The split-flow process configuration is used when the $H_2S$ concentration in the feed is between 20 and 50 mole percent. If the feed stream is leaner in $H_2S$ than 20 mole percent and at ambient temperature, the flame is not self sustaining and the feed stream must be preheated to complete the Claus reaction when the split-flow scheme is employed. The process may still be a net heat producer. Typically, the sulfur-recycle process is used when the acid gas feed contains less than 10 percent $H_2S$ by volume and the problem of a self sustaining flame occurs. These ranges of $H_2S$ concentrations are not sharply defined when used to determine which Claus process configuration should be applied in a specific situation. Factors such as the presence of compounds other than $H_2S$ and $CO_2$ in the acid gas, the acid gas feed flow rate, the stability of the acid gas composition and flow rate, past experiences of Claus plant designers, economics, and volume percent ratio of $H_2S$ and $CO_2$ have a large influence on the choice of process configuration.[4]

### 3.2.2 Claus Tail Gas Cleanup Processes

Tail gas cleanup systems can be employed to process the unconverted sulfur compounds in Claus plant tail gas. By using these systems overall sulfur recovery is increased, and $SO_2$ emissions are reduced. Even after condensation of product sulfur, the tail gases from Claus units contain appreciable quantities of $H_2S$, $SO_2$, and other sulfurous compounds. Before 1970, tail gases from Claus plants commonly were incinerated, and the resulting $SO_2$-containing gases vented through a stack to the atmoshpere. Because of the regulatory emphasis since 1970 on reducing $SO_2$ emissions, there are now several viable commercial processes available for Claus tail gas cleanup. Nearly all these cleanup processes will reduce $SO_2$ emissions to between 0.0412 and 0.0002 lb of $SO_2$ per pound of sulfur processed (97.94 to 99.99 percent) and are capable of boosting the overall sulfur recovery efficiency of a Claus plant from 93.01 percent to greater than 97.94 percent.[5]

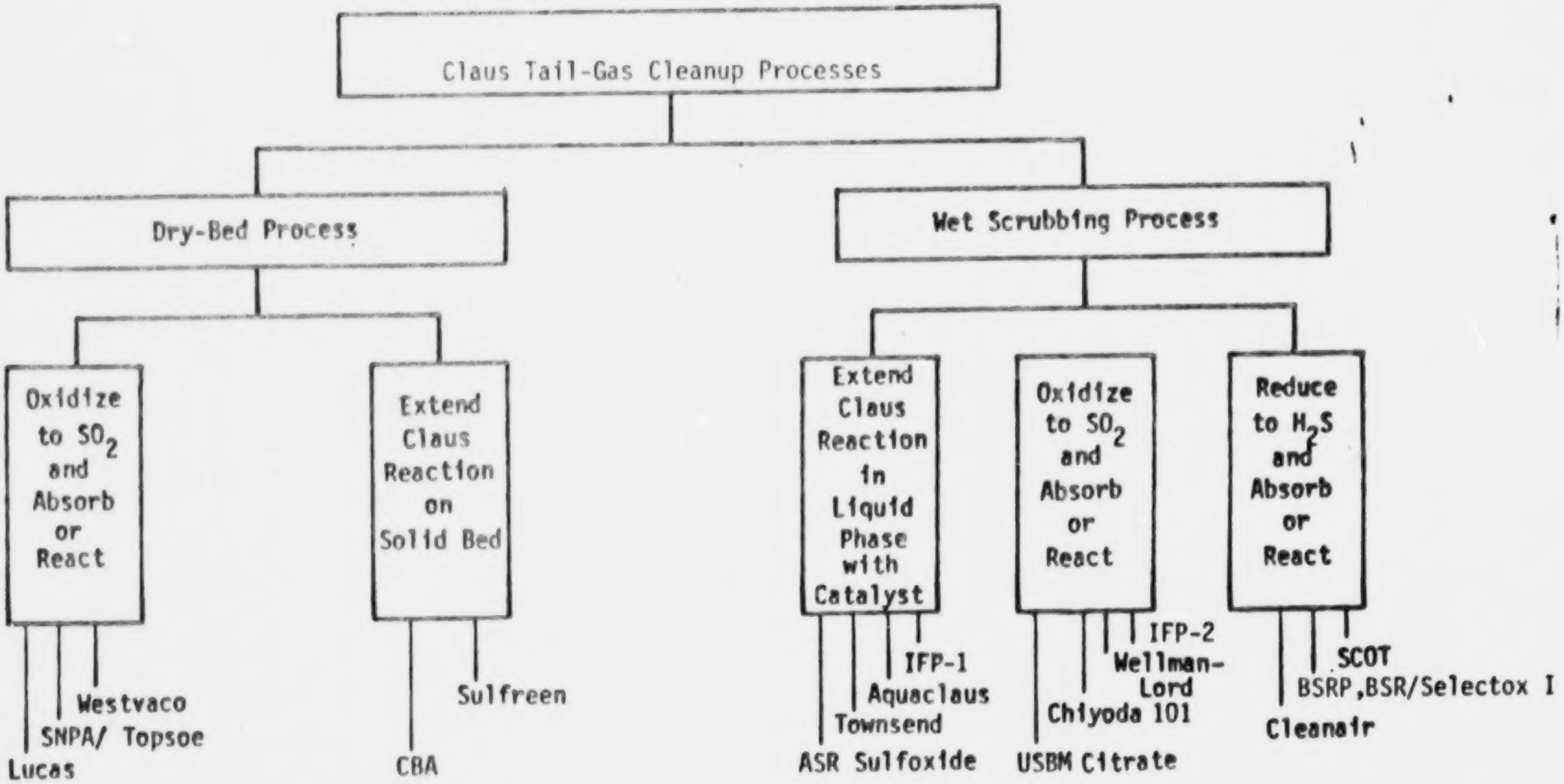
Three tail gas cleanup processes developed for reducing $SO_2$ emissions through extension of the Claus reaction by operation at low temperatures are the Cold Bed Adsorption (CBA), the IFP-1 (Institut Francais du Petrole), and the Sulfreen processes. Four other processes, the BSRP (Beavon Sulfur Removal Process), the BSR/Selectox I (Beavon Sulfur Removal/Selectox I), the SCOT (Shell Claus Offgas Treatment), and the Cleanair processes have been developed and demonstrated to reduce emissions from Claus sulfur recovery plants through conversion of the sulfur compounds present in the tail gas to $H_2S$, followed by $H_2S$ recovery. In addition, the Wellman-Lord and the IFP-2 processes were developed to reduce Claus plant tail gas sulfur emissions by conversion of sulfur compounds present in the tail gas to $SO_2$, followed by $SO_2$ recovery. The ammonium bisulfite/ammonium thiosulfate and the MCRC (Mineral and Chemical Resource Co.) limestone slurry sulfur recovery processes have also been developed. All these tail gas cleanup processes are described in detail in Chapter 4. Figure 3-3 presents these tail gas cleanup processes and other potentially available processes.

## 3.3 BASELINE CONTROL EMISSIONS LEVELS

Emissions of sulfur from onshore natural gas processing operations originate from $H_2S$ contained in "sour" natural gas deposits. This sulfur is emitted as $H_2S$ (trace amounts) and $SO_2$, with the majority of sulfur emissions occurring as $SO_2$. The major sulfur emissions sources within onshore natural gas processing operations are as follows:

1. $SO_2$ from incinerated acid gas emitted from sweetening operations,
2. $SO_2$ from incinerated tail gas emitted from sulfur recovery operations, and
3. $SO_2$ from incinerated residual tail gas emitted from tail gas cleanup systems on sulfur recovery plants.

Nationwide sulfur dioxide emissions from the existing onshore natural gas sweetening and sulfur recovery facilities in the natural gas production industry are estimated to be 250,000 megagrams per year. These


Claus Tail-Gas Cleanup Processes
Dry-Bed Process
Wet Scrubbing Process
Oxidize to SO2 and Absorb or React
Extend Claus Reaction on Solid Bed
Extend Claus Reaction in Liquid Phase with Catalyst
Oxidize to SO2 and Absorb or React
Reduce to H2S and Absorb or React
Westvaco
SNPA/ Topsoe
Lucas
Sulfreen
CBA
IFP-1
Aquaclaus
Townsend
ASR Sulfoxide
IFP-2
Wellman-Lord
Chiyoda 101
USBM Citrate
SCOT
BSRP,BSR/Selectox I
Cleanair

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CBA = Cold Bed Adsorption
BSRP = Beavon Sulfur Removal Process
IFP = Institut Francais du Petrole
SCOT = Shell Claus Offgas Treatment
SNPA = Societe Nationale des Petroles d'Aquitaine
USBM = U.S. Bureau of Mines

Figure 3-3. Claus tail gas cleanup processes.

estimates are based on the data available on 82 sulfur recovery facilities (89 facilities listed) presented in Standford Research Institute (SRI) International's Chemical Economics Handbook, December 1979 (refer to Table 9-5 and the Docket Entry A-80-20-A, II-I-44). The estimates exclude those onshore natural gas processing facilities that do not recover sulfur but simply incinerate the acid gases from the sweetening units.

### 3.3.1 Sweetening Operation Emissions

Sulfur emissions from sweetening systems of sour natural gas plants usually occur as either $H_2S$ or $SO_2$. Nearly all "sour" natural gas is routed to a gas sweetening facility where the $H_2S$ is removed and becomes part of the acid gas stream. At plants where the quantities of sulfur available in the feed gas are considered too low to be economically feasible to recover, the acid gas is often not treated for sulfur recovery. In fact, the cost would prohibit the constructing and operating of sulfur recovery units in such situations. In such cases, the sulfur in the acid gas is either vented to the atmosphere as $H_2S$ or burned with fuel gas and air in a flare or an incinerator and emitted to the atmosphere as $SO_2$. However, the $H_2S$ in acid gas is most often converted to $SO_2$ before being emitted to the atmosphere, and the fraction of $H_2S$ emissions is therefore very small compared to the total sulfur emissions in the United States. Several factors, such as the $H_2S$ concentration in the gas, state regulations, and population densities affect whether the $H_2S$ in acid gas is converted to $SO_2$ before being emitted to the atmosphere.

In addition, nitrogen oxides ($NO_x$), hydrocarbons (HC), and particulate matter may be emitted during combustion of acid gas with fuel gas. However, incinerators and flares at gas sweetening plants operate at temperatures ranging from approximately 538°C to 704°C (1,000°F to 1,300°F). At these temperatures, the combustion of $H_{2S}$ to $SO^2$ is over 99 percent complete, and significant amounts of nitrogen oxides are not formed in this low temperature range. Particulate matter and hydrocarbons emitted from sweetening operations result from incomplete combustion of the acid gas. Incomplete combustion can be caused by insufficient fuel value in the gas mixture or inadequate mixing of fuel and air. These factors are considered in incinerator and flare design and operating procedures to insure complete oxidation of $H_2S$ and other reduced sulfur compounds to $SO_2$. Therefore, smokeless flares and incinerators usually do not produce significant amounts of particulate or hydrocarbons.

### 3.3.2 Sulfur Recovery Operation Emissions

Nearly all sulfur originating in sour natural gas is extracted during sweetening operations and, if not emitted to the atmosphere after sweetening, is then routed in the acid gas to a Claus sulfur recovery plant. Unless a tail gas cleanup unit is applied, the exhaust gas leaving the Claus plant is incinerated, and consequently the tail gas sulfur is emitted to the atmosphere primarily as $SO_2$.

As in the case of burning acid gas, nitrogen oxides, hydrocarbons, and particulate matter may be emitted from Claus tail gas incineration. However, these emissions normally would be insignificant for the same reasons cited for acid gas flaring or incineration.

Sulfur compounds emissions from sulfur recovery operations are influenced by several parameters, including Claus plant reactor bed temperatures, age and condition of the catalyst used in the Claus process, the $H_2S$ and $CO_2$ concentrations in the acid gas feed, the number of conversion stages in the Claus plant, and the percent capacity at which the Claus plant is operated.

### 3.3.3 Baseline Control Emission Levels

3.3.3.1 $SO_2$ Baseline Control Emission Levels. The $SO_2$ baseline emission level is the level of emission control (1982 State $SO_2$ controls) that would exist in the natural gas production industry in the absence of any additional EPA standards. This baseline emission level is established to facilitate comparison of the economic, energy and environmental impacts of the regulatory alternatives. Current control technologies that represent industry practice are described in Chapter 4. Achievable emission reduction levels from the application of these current control technologies are described briefly in Chapter 7. Current industry practice, data from literature and vendors, and existing State requirements for sulfur recovery from natural gas production activities provide the basis for selection of the $SO_2$ baseline control emission levels.

Current technologies that represent industry practice can be classified as (1) sweetening operations with incineration of acid gases, and (2) sweetening operations followed by sulfur recovery with incineration of Claus tail gases. Assessment of actual industry practice includes considerations for sulfur recovery plant size, $H_2S$ and $CO_2$ concentrations in the acid gas feed, and number of Claus unit catalytic stages. Three baseline controls have been developed for the natural gas production industry based upon current industry practice. These baseline controls, which are applied following sweetening operations, are:

- Incineration
- Sulfur recovery and incineration (Claus-2 stage)
- Sulfur recovery and incineration (Claus-3 stage)

These baseline controls form the basis of the model plants for which cost analysis and economic impact study will be conducted. While incineration does not reduce $SO_2$ emissions, it represents a no control option, and as such is included in this discussion of control technology options. Current control technologies demonstrated at the existing facilities for efficient, economical, and continuous reduction of $SO_2$ emission levels under normal operating conditions are represented through these baseline controls.

Baseline controls and baseline emission levels for $SO_2$ are presented in Table 3-2. An assessment of the baseline control is also presented in Table 3-2, which shows sulfur recovery efficiency under normal operating conditions for a specified acid gas $H_2S/CO_2$ volume percent ratio.

Specific emission regulations for sulfur recovery plant facilities currently are applied at the State level. Direct interface with national regulations is usually confined to diffusion modeling to show compliance with ambient standards and significant deterioration standards. The degree to which applicable State regulations require $SO_2$ emission reductions depends on the size of the sulfur recovery plant and the $H_2S$ and $CO_2$ concentrations in the acid gas feed stream to the Claus unit. Many states have specific emissions standards for sulfur-recovery plants, while some use a case-by-case approach based on meeting ambient standards.[6]

3.3.3.2 $H_2S$ Baseline Control Emission Levels. Current practice of the natural gas production industry is to flare or incinerate unprocessed

Table 3-2. BASELINE CONTROLS AND BASELINE EMISSION LEVELS

| Baseline controls (no NSPS) | Size (sulfur feed rate), Mg/d | Acid gas feed $H_2S/CO_2$ ratio by volume | Baseline emission levels,[a] $SO_2$ emissions, $10^3$ Mg/y (percent average-of-run sulfur recovery) |
|---|---|---|---|
| Incinerator | $\leq 5.1$ | [b] | 3.57<br>(0.00) |
| Claus-2 stage | 10.2 | [b] | 0.62<br>(91.30)[c] |
| Claus-3 stage | 102 | [b] | 4.72<br>(93.39)[c] |
| Claus-3 stage | 564 | 50/50 | 15.39<br>(96.10) |
| Claus-3 stage | 564 | 80/20 | 13.27<br>(96.64) |
| Claus-3 stage | 1,016 | 50/50 | 27.74<br>(96.10) |
| Claus-3 stage | 1,016 | 80/20 | 23.89<br>(96.64) |

[a] 350 days/year operation.

[b] Covers the entire ratio range from less than 5/95 to over 80/20.

[c] Figures based upon 12.5/87.5 ratio.

acid gases, Claus unit tail gases, and residual tail gases (from tail gas cleanup systems) before releasing them into the ambient atmosphere. It is assumed that during thermal oxidation, operated at or above 538°C (1,000°F), all sulfur compounds, including $H_2S$, COS, and $CS_2$, are converted to $SO_2$. Therefore, separate regulations for $H_2S$ emissions from the gas processing industry have not been established, because these emissions are considered negligible in total amount.[6] However, any significant emissions of $H_2S$ from the natural gas production industry is not permitted.

## 3.4 REFERENCES FOR CHAPTER 3

1. Railroad Commission of Texas, Oil and Gas Division. Rules Having Statewide Application to Oil, Gas, and Geothermal Resource Operations within the State of Texas. Austin, Texas. October 22, 1979. Docket No. A-80-20-A, Entry II-I-43.

2. U.S. Environmental Protection Agency. Atmospheric Emissions Survey of the Sour Gas Processing Industry. Publication No. EPA-450/3-75-076. Research Triangle Park, North Carolina. October 1975. Docket No. A-80-20-A, Entry II-A-4.

3. Kohl, Arthur L. and Fred C. Riesenfeld. Gas Purification. Gulf Publishing Company. Houston, Texas. Third Edition. 1979, pp. 33-39. Docket No. A-80-20-A, Entry II-I-71.

4. Paskall, Harold G. Capability of the Modified-Claus Process. Western Research & Development. Alberta, Canada. March 1979. Docket No. A-80-20-A, Entry II-I-41.

5. The Ralph M. Parsons Company. Engineers/Constructors. Sulfur Recovery Study – Onshore Sour Gas Production Facilities – for TRW Environmental Engineering Division. July 1981. The study is presented in Appendix E. Docket No. A-80-20-A, Entry II-A-16. Also refer to Entry II-A-20 for small sulfur recovery unit study (presented in Appendix H) detailing costs for Recycle Selectox sulfur recovery facilities.

6. State Air Laws. Environmental Reporter. The Bureau of National Affairs, Inc. Washington, D.C. 1979-1981. Docket No. A-80-20-A, Entry II-I-65.

# 4. EMISSION CONTROL TECHNIQUES

## 4.1 GENERAL DESCRIPTION

For many years the Claus Sulfur Recovery Process has been utilized to convert hydrogen sulfide ($H_2S$) recovered from sour natural gas treating processes into elemental sulfur. This process is the first major step in reducing the amount of this noxious gas that reaches the atmosphere. Practically all of the sulfur recovery plants in this country are based on some version of the Claus process for recovering sulfur from $H_2S$.

Since 1938, the principal improvements to the Claus process have been obtained by sequential addition of more catalytic reactors, with sulfur and heat removal between the reactors to shift the equilibrium of the Claus reaction towards higher sulfur removals. This process has now advanced to the point that sulfur recovery efficiencies have increased from the 90 to 92 percent levels first obtained in the late 1940's to present recovery efficiency levels of up to 97 percent of inlet sulfur.[1,2] Numerous processes have also been developed during the past decade to remove residual sulfur compounds from Claus plant tail gas. Several of these tail gas cleanup processes are designed to boost the sulfur recovery capability of the Claus system to 99.9 percent recovery of sulfur contained in the acid gas feed to the Claus plant.[3]

In general, the sulfur compound emission conversion or control techniques utilized in the natural gas production industry consist of incinerators and flaring units, two- and three-stage Recycle Selectox sulfur recovery units, two- and three-stage Claus sulfur recovery units, and a variety of Claus tail gas cleanup systems. The sulfur recovery technologies currently used in the natural gas production industry are available to control sulfur compounds emissions from acid gas streams over the entire range of $H_2S$ concentration. However, factors such as the $H_2S$ and carbon dioxide ($CO_2$) concentrations in the acid gas feed, the expected throughput (sulfur intake) of the facility, the location, and economics greatly influence the Claus process configuration, the number of Claus reactors, the type of tail gas cleanup unit, if any, and Recycle Selectox process, selected for a particular sulfur recovery facility.

## 4.2 SULFUR EMISSION CONTROL TECHNOLOGIES USED IN THE INDUSTRY

### 4.2.1 Incineration

Incineration is a temperature-controlled process in which fuel gas and air are supplied to the incinerator in sufficient quantities to maintain a constant temperature of 538°C - 704°C (1,000°F - 1,300°F) to ensure complete oxidation of $H_2S$ and other reduced sulfur compounds, such as carbon disulfide ($CS_2$) and carbonyl sulfide (COS), to sulfur dioxide ($SO_2$). Employing incineration does not reduce the total amount of sulfur compound emissions, however, the technique does convert $H_2S$ into less toxic $SO_2$. If the $H_2S$ concentration in the acid gas discharged from a natural gas sweetening facility is very low or the acid gas throughput is small, the acid gas usually is incinerated and the resultant $SO_2$ discharged into the atmosphere. Incineration is always employed as the final step in the Claus sulfur recovery process (with no tail gas cleanup unit). If a tail gas cleanup unit is applied to the Claus plant, the residual tail gas from the cleanup unit is generally incinerated. However, incineration normally is not used with the Beavon Sulfur Removal Process (BSRP). In this process the incinerator is kept as a stand-by unit and used only when necessary.

### 4.2.2 2-Stage Claus Sulfur Recovery Process

A Claus sulfur recovery facility utilizing two catalytic stages generally is capable of attaining recovery design efficiencies of up to 96.31 percent, depending on the $H_2S$ concentration in the acid gas fed to the Claus unit. If the $H_2S$ concentration in the acid feed gas is as low as 12.5 percent by volume, however, a two-stage Claus plant may be able to achieve only a 93.01 percent sulfur recovery design efficiency.[4] These efficiency levels are achieved when the catalysts are fresh. Due to sulfation, carbon deposition, physical attrition among other reasons, the Claus catalysts (alumina) gradually degrade with time. Sulfur recovery design data indicates that catalyst degradation results in an average of approximately 1.14 percent reduction in efficiency per year for a Claus 2-stage unit.[4] The data also indicates that average-of-run efficiency is an average of the efficiency when the catalysts are fresh and the efficiency when the catalysts are spent and about to be replaced.[4]

For Claus plants having two catalytic converter stages, the following sulfur recovery efficiencies have been reported.[4]

| $H_2S/CO_2$ mole percent ratio in acid gas feed | Sulfur recovery efficiency | |
|---|---|---|
| | Start-of-run (fresh catalyst) | End-of-run (spent catalyst) |
| 12.5/87.5 | 93.01 | 89.59 |
| 20/80 | 93.46 | 90.14 |
| 50/50 | 94.99 | 92.61 |
| 80/20 | 96.31 | 93.89 |

A flow diagram for a typical two-stage Claus sulfur recovery plant is shown in Figure 4-1. The first step in the Claus process is the complete oxidation of one-third of the $H_2S$ in the acid gas to $SO_2$ (the thermal phase):

$$3H_2S + \frac{3}{2}O_2 \xrightarrow[1090^\circ C]{Flame} SO_2 + H_2O + 2H_2S + 520.9\ kJ \qquad (4\text{-}1)$$

This step is carried out in a reaction furnace, and the released energy is recovered and used to generate steam. In the second step (the catalytic phase), the remaining two-thirds of the $H_2S$ is reacted over a catalyst with the $SO_2$ produced in the initial reaction:

$$2H_2S + SO_2 \xrightarrow{catalyst} 3S + 2H_2O + 92.9\ kJ \qquad (4\text{-}2)$$

The flow diagram shown in Figure 4-1 indicates that the Claus process can be separated into two phases, a thermal phase and a catalytic phase. In the thermal phase, the temperature in the reaction furnace is usually between 980°C and 1090°C (1800°F and 2000°F). During the catalytic phase, the temperature is maintained below 370°C (700°F) and somewhat

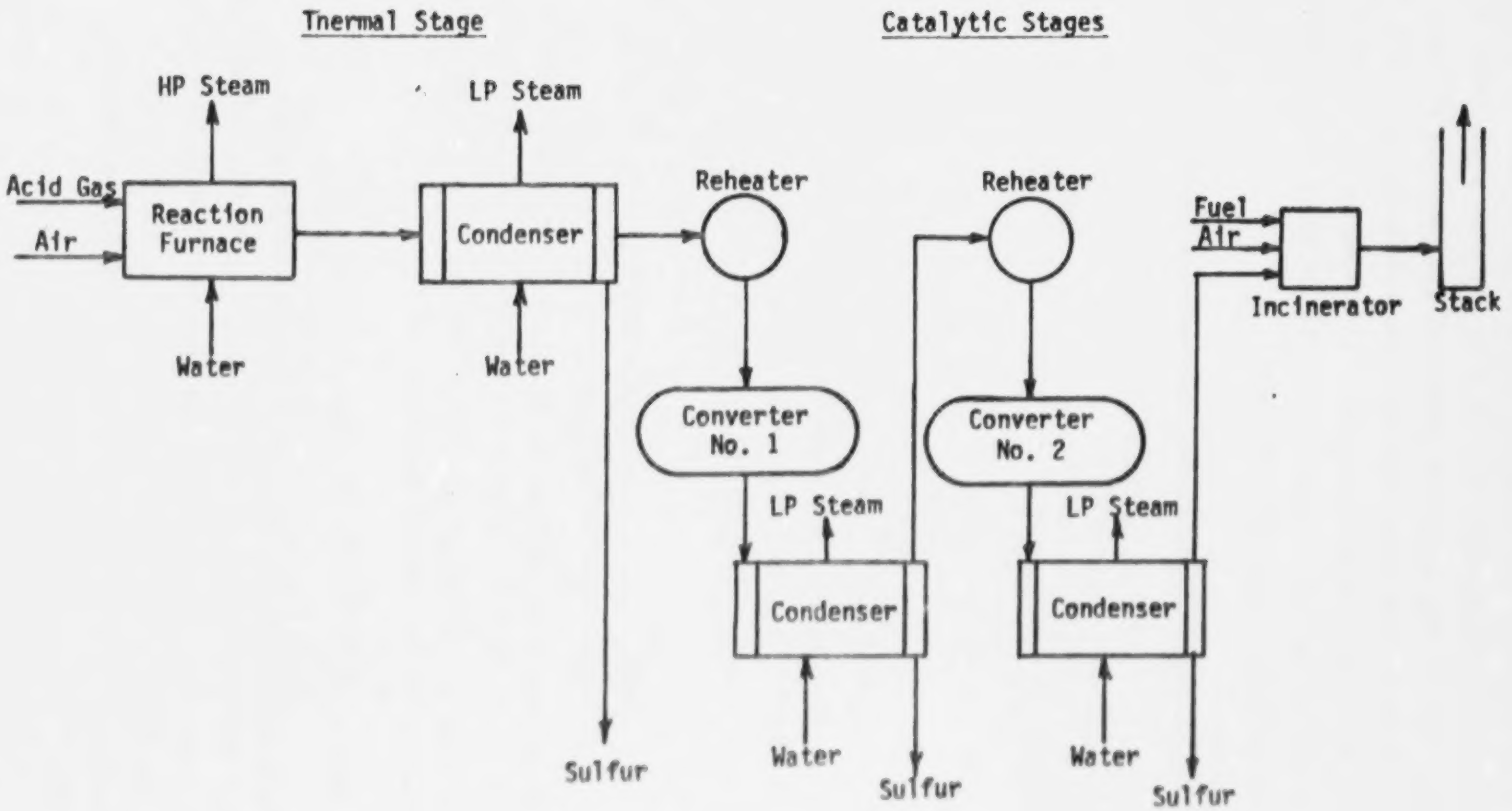
Thermal Stage
Catalytic Stages
HP Steam
LP Steam
Acid Gas
Air
Reaction Furnace
Condenser
Water
Water
Reheater
Reheater
Fuel
Air
Incinerator
Stack
Converter No. 1
Converter No. 2
LP Steam
LP Steam
Condenser
Condenser
Sulfur
Water
Sulfur
Water
Sulfur

Figure 4-1. Flow diagram for a two-stage Claus sulfur recovery plant.

above the sulfur dew point of the gas mixture. The normal operating temperature during the catalytic phase is between 250°C and 300°C (475°F and 575°F), but the lower this temperature, the more complete is the conversion to sulfur. Therefore, it is advantageous to provide several catalytic stages with condensation of the sulfur formed after each stage. After the sulfur is condensed following each Claus converter, the process gas is preheated to reaction temperature before entering the next catalytic reactor. The economic incentive of high sulfur prices and continued demand for liquid sulfur induce an increasing number of producers to recover sulfur from $H_2S$.[5]

### 4.2.3 3-Stage Claus Sulfur Recovery Process

Utilizing three catalytic stages, a typical Claus sulfur recovery facility is capable of attaining efficiencies of 94.72 to 97.31 percent of the sulfur in the acid gas stream under normal conditions. These efficiency levels are achieved when the catalysts are fresh. Due to sulfation, carbon deposition, physical attrition among other reasons, the Claus catalysts (alumina) gradually degrade with time. Sulfur recovery design data indicates that catalyst degradation results in an average of approximately 0.89 percent reduction in efficiency per year for a Claus 3-stage unit.[4] The data also indicate that average-of-run efficiency is an average of the efficiency when the catalysts are fresh and the efficiency when the catalysts are spent and about to be replaced.[4]

For Claus plants having three catalytic converter stages, the following sulfur recovery efficiencies have been reported.[4]

| $H_2S/CO_2$ mole percent ratio in acid gas feed | Sulfur recovery design efficiency[4] |  | Sulfur recovery actual efficiency (from API)[12] |
|---|---|---|---|
|  | Start-of-run (fresh catalyst) | End-of-run (spent catalyst) |  |
| 12.5/87.5 | 94.72 | 92.06 | 93.7 |
| 20/80 | 95.60 | 93.40 | 94.1 |
| 50/50 | 96.85 | 95.35 | 95.2 |
| 80/20 | 97.31 | 95.97 | 96.2 |

The sulfur recovery efficiency of Claus plants decreases with a decrease in $H_2S$ concentration. The loss in sulfur recovery efficiency is associated with increased concentrations of $CO_2$ in the acid gas feed because the Claus process is equilibrium-limited, and increased concentrations of $CO_2$ and impurities drive the process away from the recovery of sulfur. These process limitations occur regardless of the $H_2S$ concentration in the acid gas feed; but they are more significant, and sulfur losses become greater as the $H_2S$ feed concentration decreases.

To improve sulfur recoveries at Claus plants where the acid gas feed stram contains a lean $H_2S$ concentration, a modified Claus process can be utilized. Four basic configurations of the modified Claus process that are being employed currently by the onshore natural gas production industry are the straight-through, split-flow, split-flow with preheating, and sulfur recycle configurations. The differences among these process configurations are in the methods used to produce the $SO_2$ prior to the first converter.

4.2.3.1 Straight-Through Configuration. If the acid gas feed to the Claus unit contains 50 mole percent or greater of $H_2S$, the straight-through configuration is normally utilized, because it provides the highest overall sulfur recovery and permits maximum heat recovery at a high temperature.[6,7]

In the straight-through process scheme, the entire acid-gas stream and the stoichiometric amount of air to burn one-third of the $H_2S$ to $SO_2$ are fed through a burner to the reaction furnace. Then, sufficient retention time is provided to allow the generated $SO_2$ to react with the unburned $H_2S$ to form sulfur vapor. At the temperatures prevailing in the reaction furnace, typically above 1090°C (2000°F), a substantial amount of elemental sulfur is formed. The elemental sulfur is condensed after the gases are cooled first in a waste heat boiler and then in a sulfur condenser. Up to 70 percent of the overall conversion of $H_2S$ to elemental sulfur can take place in this thermal-conversion step. Although high-pressure steam can be generated in the waste heat boiler, it is preferable to produce low-pressure steam (15 to 50 psig) in the sulfur condenser in order to cool the reaction gases to obtain maximum sulfur condensation. Many designs often use the heat from the fourth condenser to preheat boiler feed water in place of generating steam. The reaction gases leaving the sulfur condenser are reheated and passed through the first catalytic converter where additional sulfur is produced by the reaction of $H_2S$ with $SO_2$. The chemical reaction equations are presented in Section 4.2.2. Although these reactions are exothermic, reheating of the gas is necessary to maintain the temperature above the sulfur dew point as the gas passes through the catalytic converter, because condensation and deposition of sulfur on the catalyst causes catalyst deactivation. The gases leaving the first catalytic converter are again cooled, and sulfur is condensed. The process of reheating, catalytically reacting, and sulfur condensing is repeated two more times in a three-stage sulfur recovery unit. As conversion progresses through the catalytic stages and more sulfur is removed from the gas mixture, the sulfur dew point of the reaction gases is lowered, permitting operation at progressively lower temperatures, thus improving conversion. After leaving the last sulfur condenser, the exhaust gases are either incinerated to convert all remaining sulfur compounds to $SO_2$ before being emitted to the atmosphere or further treated in a separate tail gas cleanup process.

Figure 4-2 shows a flow diagram for a three-stage Claus plant utilizing the straight-through process configuration.

4.2.3.2 Split-Flow Configuration. The split-flow process is used for acid gas streams containing $H_2S$ in such low concentrations that stable combustion could not be sustained if the entire gas stream were fed to the reaction furnace. Generally, the split-flow configuration is employed at Claus sulfur recovery facilities where the $H_2S$ concentration in the acid gas feed stream is between 20 and 50 mole percent.[7]

In the split-flow process, one-third of the acid gas is fed to the reaction furnace, and all the $H_2S$ contained in the gas is combusted with the stoichiometric amount of air to form $SO_2$. In this configuration, most of the $H_2S$ fed to the furnace is oxidized to $SO_2$, and little or no sulfur is produced in the furnace. Then, the hot gases are cooled in a waste heat boiler and combined with the remaining two-thirds of the acid gas before entering the first catalytic conversion stage. The Claus reactor and condenser train of the split-flow process configuration is very similar or identical to that of the straight-through process.

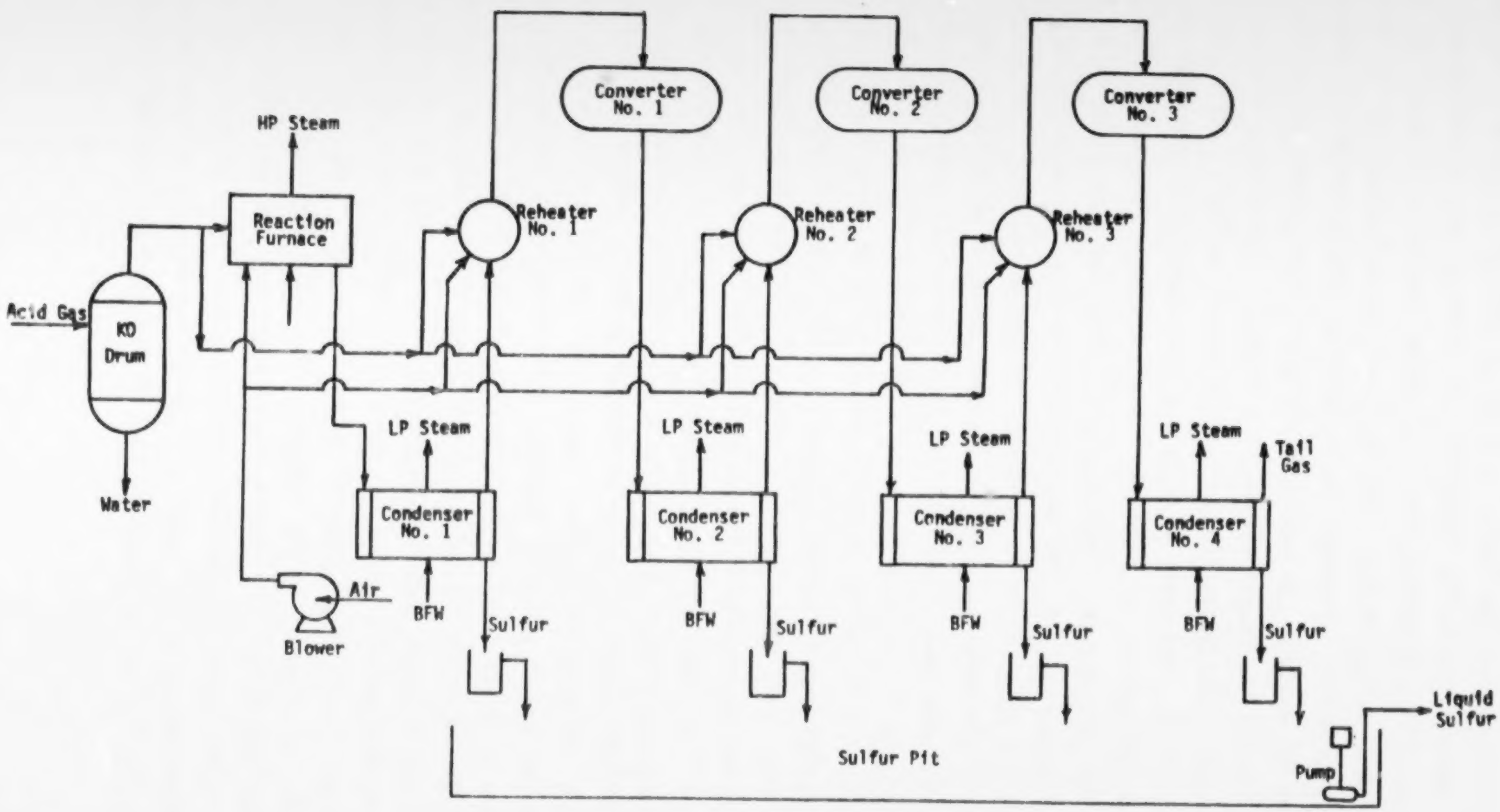
Acid Gas
KO Drum
Water
HP Steam
Reaction Furnace
Air
Blower
Reheater No. 1
Converter No. 1
Reheater No. 2
Converter No. 2
Reheater No. 3
Converter No. 3
LP Steam
Condenser No. 1
BFW
Sulfur
LP Steam
Condenser No. 2
BFW
Sulfur
LP Steam
Condenser No. 3
BFW
Sulfur
LP Steam
Tail Gas
Condenser No. 4
BFW
Sulfur
Sulfur Pit
Pump
Liquid Sulfur

Figure 4-2. Flow diagram for a three-stage Claus sulfur recovery facility utilizing the straight-through process configuration.

Figure 4-3 shows a flow diagram for the split-flow configuration of the modified Claus process.

4.2.3.3 Split-Flow With Preheating of Feed Streams Configuration. In cases where the $H_2S$ concentration in the acid gas feed is too low to achieve a sufficiently high temperature in the reaction furnace using the two-third bypass employed in the split-flow configuration, acid gas preheating or the sulfur-recycle configuration is often utilized. Acid gas preheating commonly is used to permit processing by the split-flow method for Claus acid gas feed streams containing 10 to 25 mole percent $H_2S$.[7] In addition, recent success has been reported on the use of the split-flow configuration with preheating of feed streams, with an acid gas feed stream containing only 8 mole percent $H_2S$ where previously the use of direct oxidation failed because of severe catalyst deactivation.[8]

In this configuration of the Claus process, adequate furnace temperature is achieved by preheating both air and acid gas and by adding a supplemental fuel gas to the flame. Problems with $CS_2$ formation are avoided by use of a specially designed burner in which a set of concentric pipes carry the fuel gas, fuel gas air, acid gas, and process air.

4.2.3.4 Sulfur Recycle Configuration. Generally, the sulfur recycle configuration of the Claus process is used when the $H_2S$ concentration in the acid gas feed to a Claus sulfur recovery facility is less than 10 mole percent.[7] In the sulfur recycle Claus process, product sulfur is recycled to the furnace and burned with air to produce $SO_2$. Then, the $SO_2$ formed in the furnace is combined with the preheated acid gas feed stream and fed to the Claus reactors as shown in Figure 4-4. The reactor-condenser train of the sulfur recycle configuration is similar to that of the straight-through configuration.

### 4.2.4 Recycle Selectox Process

The Selectox catalyst in this process enables $H_2S$ to be oxidized to sulfur with air at a low temperature without forming $SO_3$ or oxidizing either hydrogen or light hydrocarbons.[13] This eliminates the need for high-temperature combustion as in the Claus sulfur recovery process. The limited test information available for the Selectox process indicates that it is not affected by the presence of other compounds and that it

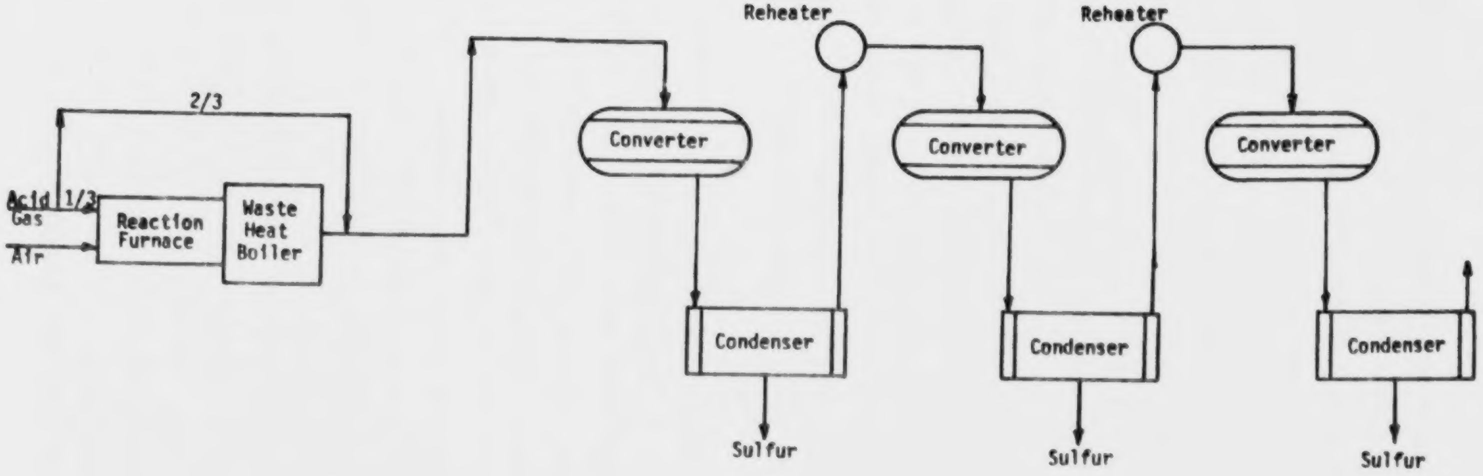
Acid Gas
1/3
Air
2/3
Reaction Furnace
Waste Heat Boiler
Converter
Condenser
Sulfur
Reheater
Converter
Condenser
Sulfur
Reheater
Converter
Condenser
Sulfur

Figure 4-3. Flow diagram for the split-flow configuration of the Claus sulfur recovery process.

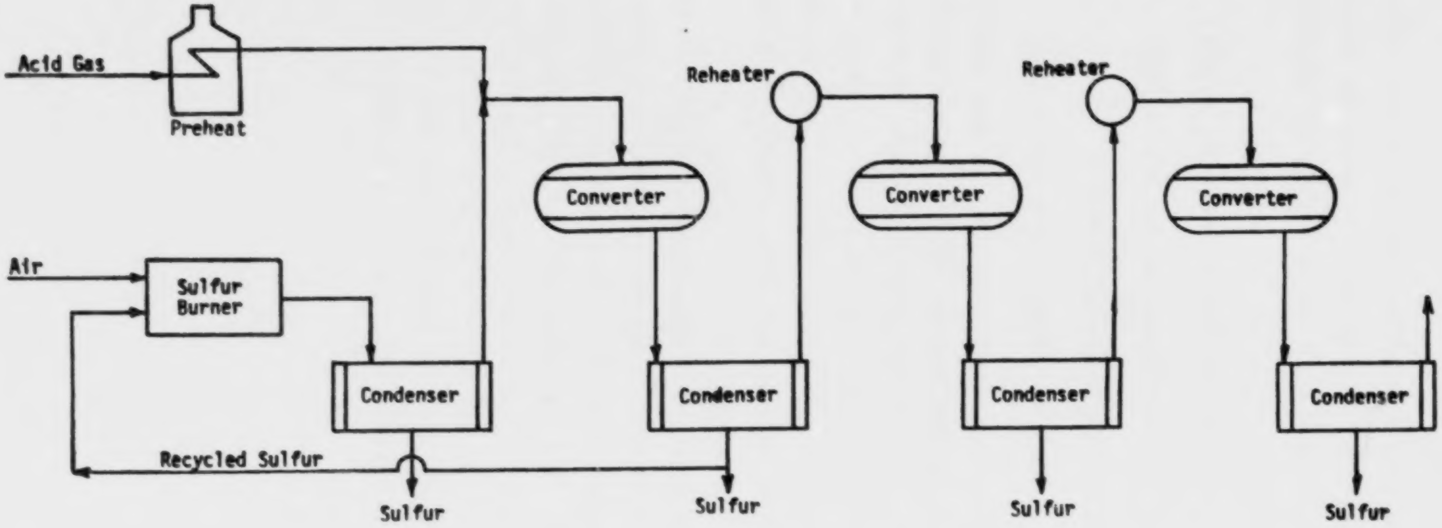
Acid Gas
Preheat
Air
Sulfur Burner
Condenser
Recycled Sulfur
Sulfur
Converter
Condenser
Sulfur
Reheater
Converter
Condenser
Sulfur
Reheater
Converter
Condenser
Sulfur

Figure 4-4. Flow diagram for the sulfur recycle configuration of the Claus sulfur recovery process.

may operate at a higher temperature than other processes without losing activity or specificity.[14]

When supplied with the proper amount of air, the oxidation of one-third of the $H_2S$ to $SO_2$ and reaction with the remaining two-thirds of $H_2S$ occur simultaneously in the presence of the Selectox catalyst to form elemental sulfur:

$$H_2S + 3/2\ O_2 \longrightarrow SO_2 + H_2O \qquad (4\text{-}3)$$

$$2H_2S + SO_2 \longrightarrow 3S + 2H_2O \qquad (4\text{-}4)$$

The effective overall reaction is therefore:

$$3H_2S + 3/2\ O_2 \longrightarrow 3S + 3H_2O \qquad (4\text{-}5)$$

The Selectox Process (once-through) is employed when the acid gas feed strength is 5 percent $H_2S$ or less. The exothermic reaction with 5 percent gas results in a reasonable maximum reactor outlet temperature. The gas exiting from the Selectox reactor is cooled in a sulfur condenser that produces steam and condenses sulfur. Gas from the sulfur condenser then proceeds through one or two Claus stages, each with a reheater, converter, and sulfur condenser.

When the acid gas feed contains more than 5 percent $H_2S$, some of the cooled lean gas from the sulfur condenser is reheated and recycled to the Selectox reactor inlet to maintain approximately 5 percent $H_2S$ at the reactor inlet. This arrangement is called Recycle Selectox process.

The Recycle Selectox 2-Stage Process, shown in Figure 4-5, has one Selectox and one Claus stage. The Recycle Selectox 3-Stage Process, illustrated in Figure 4-6, is identical but has a second Claus stage added. Achievable design sulfur recovery efficiencies extracted from Appendix H are:

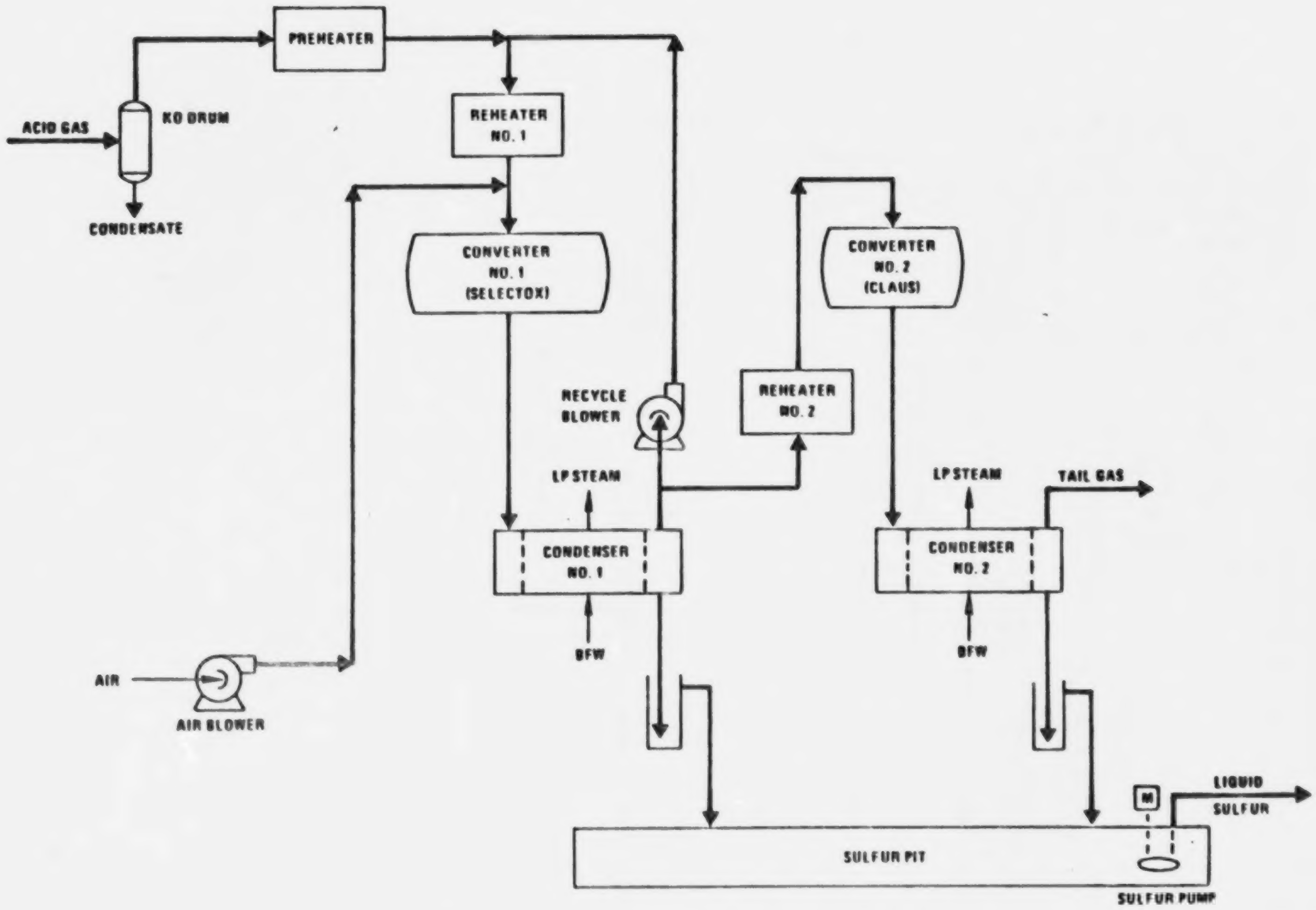
ACID GAS
KO DRUM
CONDENSATE
PREHEATER
REHEATER NO. 1
CONVERTER NO. 1 (SELECTOX)
AIR
AIR BLOWER
RECYCLE BLOWER
LP STEAM
CONDENSER NO. 1
BFW
REHEATER NO. 2
CONVERTER NO. 2 (CLAUS)
LP STEAM
TAIL GAS
CONDENSER NO. 2
BFW
M
LIQUID SULFUR
SULFUR PIT
SULFUR PUMP

Figure 4-5. Recycle Selectox 2-Stage Process.

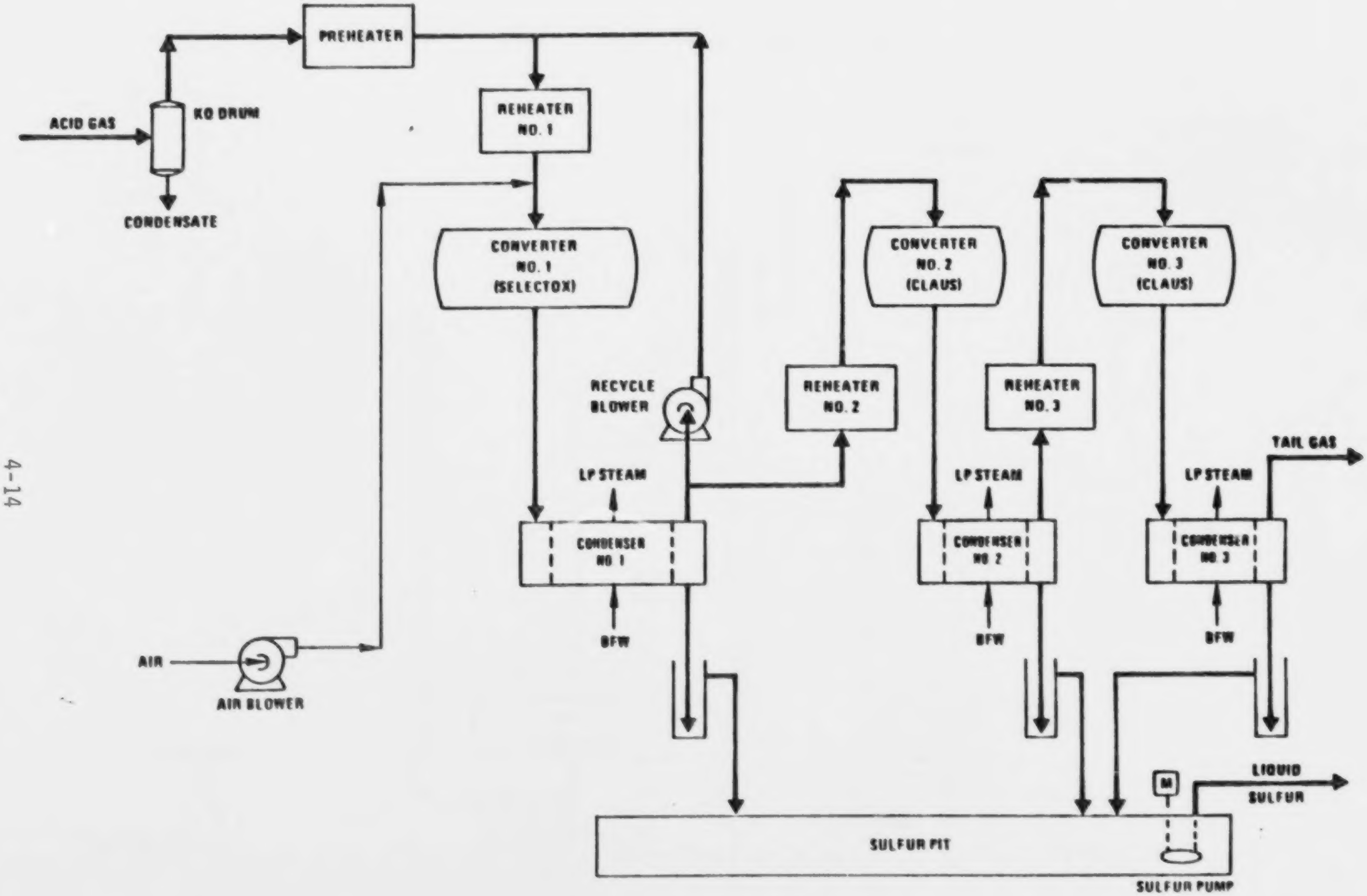
ACID GAS
KO DRUM
CONDENSATE
PREHEATER
REHEATER NO. 1
CONVERTER NO. 1 (SELECTOX)
AIR
AIR BLOWER
RECYCLE BLOWER
LP STEAM
CONDENSER NO. 1
BFW
REHEATER NO. 2
CONVERTER NO. 2 (CLAUS)
LP STEAM
CONDENSER NO. 2
BFW
REHEATER NO. 3
CONVERTER NO. 3 (CLAUS)
LP STEAM
CONDENSER NO. 3
BFW
TAIL GAS
M
LIQUID SULFUR
SULFUR PIT
SULFUR PUMP

Figure 4-6. Recycle Selectox 3-Stage Process.

| Process | $H_2S/CO_2$ ratio | Percent efficiency Start-of-run (fresh catalyst) | End-of-run spent catalyst) |
|---|---|---|---|
| Selectox 2-stage | 2/98 | 80.68 | 74.83 |
| | 5/95 | 87.90 | 81.65 |
| | 12.5/87.5 | 92.36 | 85.66 |
| Selectox 3-stage | 2/98 | 83.61 | 76.92 |
| | 5/95 | 91.40 | 84.00 |
| | 12.5/87.5 | 95.12 | 87.60 |

The sulfur recovery design data presented in Appendix H indicate that catalyst degradation (assuming 4 year average of the catalyst's life) results in an average of approximately 1.68 percent reduction in efficiency per year for a Recycle-Selectox 2-stage unit and an average of approximately 1.88 percent reduction in efficiency per year for a Recycle Selectox 3-stage unit.

The Selectox and the Claus reactor system are the same: heating gas to the desired inlet temperature, reaction in a converter, and cooling the gas and condensing sulfur in a condenser. The condensers produce low-pressure steam.

### 4.2.5 3-Stage Claus Unit With Tail Gas Cleanup Processes

The trend to reduce emissions of sulfur compounds to the atmosphere because of increasingly stringent air pollution regulations has created the need for a new group of processes designed to clean up the tail gas from Claus sulfur recovery units. There are four commercially available tail gas cleanup processes which represent a wide spectrum of process technologies and capabilities of reducing $SO_2$ emissions to the atmosphere. These are SCOT, BSRP, Sulfreen and BSR/Selectox I processes. Several other processes are at different stages of development.[9] Prior to the selection and application of a particular process technology, a detailed review of all commercially viable processes should be made to determine the most economical or optimum tail gas cleanup process scheme to be used for reducing $SO_2$ emissions from new or existing Claus sulfur recovery plants.

The four most widely used tail gas cleanup processes are discussed in the following subsections. In addition, several other viable processes for removing sulfur compounds from Claus plant tail gas are described briefly in Subsection 4.2.5.5 of this chapter. It should be noted that a tail gas cleanup unit can also be used with a 2-stage Claus unit and that the number of stages is an economics decision.

4.2.5.1 3-Stage Claus Unit With SCOT. The Shell Claus offgas treating (SCOT) process, announced to industry in 1972, was developed by the Royal Dutch/Shell Laboratories in the Netherlands and is licensed in the United States by the Shell Development Company in Houston. The SCOT process is capable of increasing the sulfur recovery efficiency of Claus units from the usual level of about 95 percent to more than 99.8 percent.[10] Sulfur recovery design data[4] indicates start-of-run efficiency (when the catalysts are fresh) with 12.5 percent $H_2S$ in acid gas is 99.89 percent and with 80 percent $H_2S$ in acid gas is 99.99 percent. However, catalysts gradually degrade with time due to sulfation, carbon deposition, physical attrition among other reasons. The data[4] indicates that catalyst degradation results in approximately 0.013 percent reduction in efficiency per year for a SCOT unit. Average-of-run efficiency is an average of the efficiency when the catalysts are fresh and the efficiency when the catalysts are spent and about to be replaced.[4]

The SCOT process consists of essentially three stages: (1) heating and reduction; (2) cooling and quenching; and (3) $H_2S$ absorption, stripping, and recycle. In the first SCOT stage, the Claus plant tail gas is heated to about 300°C (570°F) and reacted with hydrogen or a mixture of hydrogen and carbon monoxide over a cobalt molybdenum catalyst. All sulfurous compounds in the tail gas, including $SO_2$, sulfur vapor (S), COS, and $CS_2$, are reduced to $H_2S$. The hot gas from these highly exothermic reactions is cooled in a waste-heat boiler and finally quenched in a water-quench tower to about ambient temperature. In the final stage, the $H_2S$ in the gas is selectively absorbed in an alkanolamine solution. The effluent gas from the SCOT absorber, containing about 200 ppmv-500 ppmv of $H_2S$ is incinerated before it is discharged to the atmosphere. The rich amine is stripped in a conventional manner, and the $H_2S$-rich stream is recycled to the front of the Claus unit. A flow diagram for the SCOT process is given in Figure 4-7.

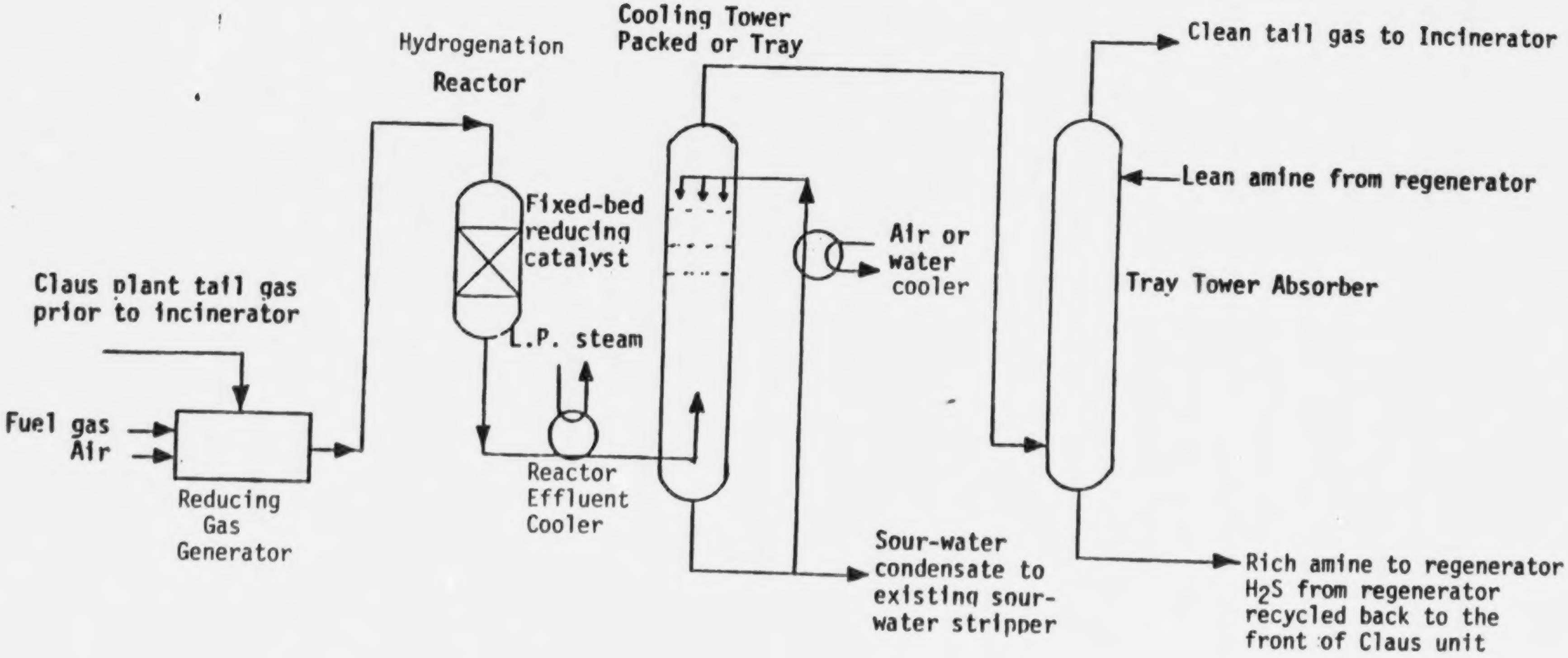
Claus plant tail gas prior to incinerator
Fuel gas
Air
Reducing Gas Generator
Hydrogenation Reactor
Fixed-bed reducing catalyst
L.P. steam
Reactor Effluent Cooler
Cooling Tower Packed or Tray
Air or water cooler
Sour-water condensate to existing sour-water stripper
Clean tail gas to Incinerator
Lean amine from regenerator
Tray Tower Absorber
Rich amine to regenerator H2S from regenerator recycled back to the front of Claus unit

Figure 4-7. Flow diagram for the Shell Claus Off-gas Treatment (SCOT) process.

Several advantages of the SCOT process are (1) easy adaption to an existing Claus plant, (2) use of familiar process technology and equipment, (3) easy and flexible operation, (4) elimination of secondary air and water pollution, and (5) high degree of sulfur removal over a wide range of operating conditions. The disadvantages are (1) higher capital cost, (2) higher fuel usage, and (3) higher heating and cooling requirements. The SCOT process, however, is considered to be one of the most flexible tail gas cleanup processes available commercially.[9,10]

Currently, more than thirty-five SCOT units are in operation worldwide on Claus plants ranging in size from 3 metric tons to 2,100 metric tons of sulfur intake per day.[3] Also, more than forty other SCOT systems are in engineering, construction, or start-up phases.[10] In onshore production activities, one SCOT unit is in operation, and two are in engineering and start-up phases.

4.2.5.2 2-Stage Claus Unit With Sulfreen Unit. Lurgi Apparante-Technik of Frankfurt, West Germany and France's Société Nationale des Pétroles d'Aquitaine (SNPA) combined their efforts to develop the Sulfreen process for Claus tail gas cleanup. The Sulfreen process, which is essentially an extension of the Claus reaction, is capable of boosting the overall sulfur recovery efficiency of the Claus/Sulfreen system up to 98.82 percent.[4] This is achieved when the catalysts are fresh. Due to sulfation, carbon deposition among other reasons, the catalysts (alumina) gradually degrade with time. Sulfur recovery design data indicates that catalyst degradation results in approximately 0.29 percent reduction per year for a Sulfreen unit.[4] The data indicates that average-of-run efficiency is an average of the efficiency when the catalysts are fresh and the efficiency when the catalysts are spent.[4] The percentage of sulfur recovery attainable by the Sulfreen unit depends upon the concentration of $H_2S$ and $SO_2$ in the Claus tail gas fed to the unit.

The Sulfreen process converts $H_2S$ and $SO_2$ contained in the tail gas to sulfur at low temperatures of 127°C to 150°C (260°F to 300°F) by extension of the classic Claus reaction:

$$2H_2S + SO_2 \xrightarrow{\text{catalyst}} 3S + 2H_2O \qquad (4\text{-}6)$$

A special activated alumina is used as an adsorbent and a catalyst in the reactors. This material was selected because of its high adsorption capacity and ease of desorption of sulfur deposited on its surfaces. Sulfur formed during the reaction is adsorbed as a liquid on the catalyst, which removes it from the reaction zone, thereby allowing the reaction to move further to the right to obtain a higher conversion than in the Claus process. This reduces entrained sulfur to a minimum. A Sulfreen unit consists of at least two parallel reactors, one in adsorption and one in desorption service. The number of reactors needed for a particular Sulfreen unit is determined strictly by economic considerations. Desorption of sulfur is accomplished by means of hot gas in a closed cycle, after a period of operation during which the catalyst has adsorbed its limit of sulfur. The hot gas is passed through the reactor to strip sulfur from the catalyst. The Sulfreen unit operation is simple and differs only slightly from that of a Claus unit. There are usually three reactors, two of which are in adsorption service while one is being regenerated. Since only solid adsorbents are used and no liquids are produced except sulfur, the process is free of liquid waste disposal problems. Sulfur that is condensed and drained to the sulfur pit is bright yellow and 99.9 percent pure and can be combined with Claus produced sulfur. A flow diagram for the Sulfreen process is presented in Figure 4-8.

The Sulfreen process is a viable, commercially proven process.[9] Currently, at least nineteen Sulfreen tail gas cleanup units are in operation worldwide and several other Sulfreen units are in construction. The Sulfreen process is quite attractive for large Claus plants, due mainly to the possibility of eliminating a fourth Claus converter and reducing the size and height of the tail gas disposal stack when the Sulfreen unit is utilized. Also, the process may be attractive for smaller Claus plants where an $SO_2$ emission level of 1,500 ppmv to 2,000 ppmv is permitted.

4.2.5.3 3-Stage Claus Unit With Beavon. The Beavon Sulfur Removal Process (BSRP) was developed by the Ralph M. Parsons Company and Union

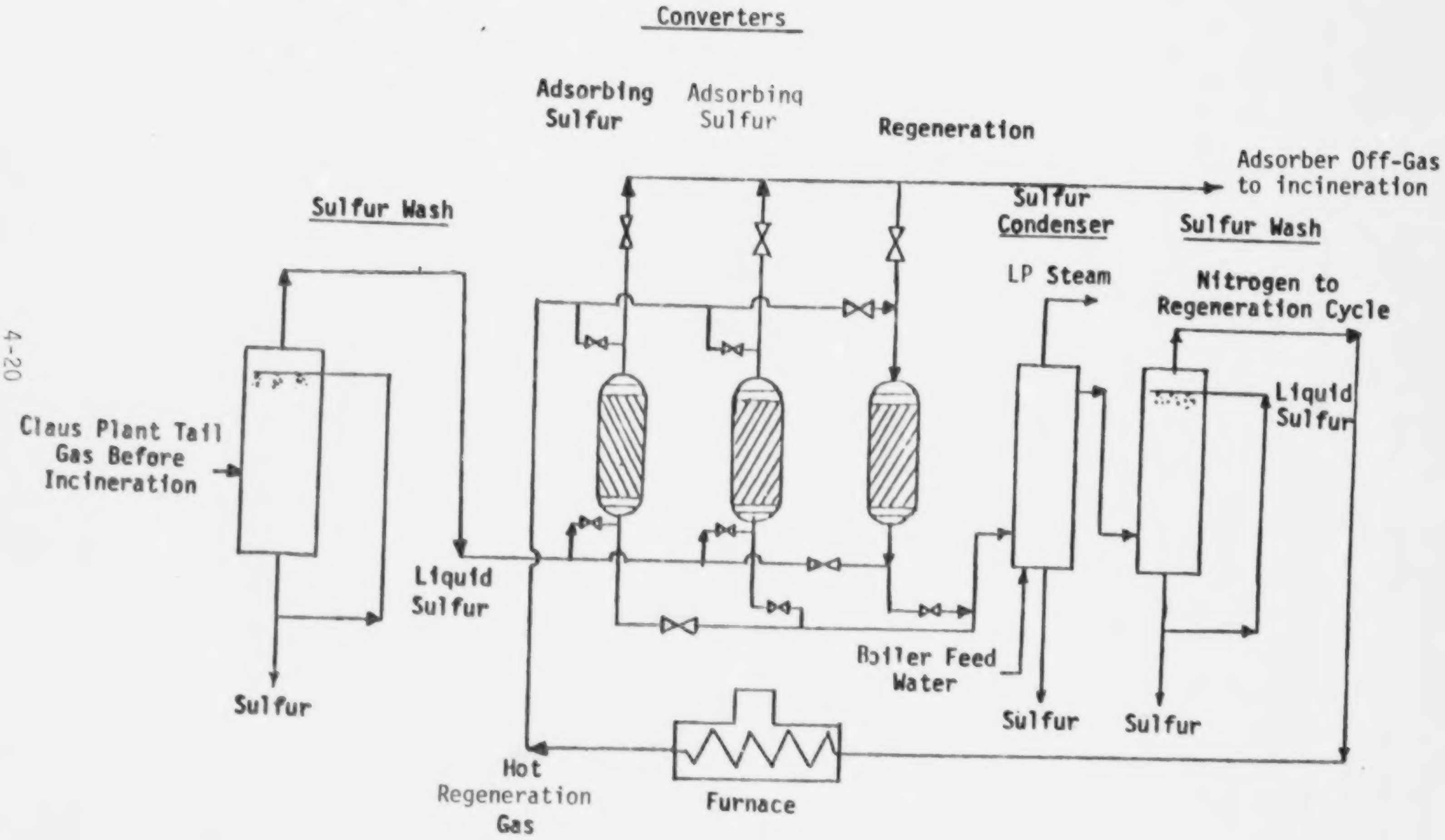
Converters
Adsorbing Sulfur
Adsorbing Sulfur
Regeneration
Adsorber Off-Gas to incineration
Sulfur Wash
Sulfur Condenser
Sulfur Wash
LP Steam
Nitrogen to Regeneration Cycle
Claus Plant Tail Gas Before Incineration
Liquid Sulfur
Liquid Sulfur
Boiler Feed Water
Sulfur
Sulfur
Sulfur
Hot Regeneration Gas
Furnace

Figure 4-8. Flow diagram for the Sulfreen tail gas cleanup process.

Oil Company of California to clean up tail gas streams from Claus sulfur recovery units. A BSRP tail gas cleanup system applied to a three-stage Claus unit can boost overall sulfur recovery to 99.99 percent (start-of-run), leaving a sulfur concentraton of 250 ppm or less in the residual tail gas.[4]

The BSRP process employs three distinct steps: (1) hydrogenation of sulfurous compounds to $H_2S$ in a catalytic converter, (2) cooling of the converter-effluent gases, and (3) conversion of the $H_2S$ in the tail gas from the cooler to elemental sulfur by use of the Stretford process. First, the Claus unit tail gas is heated to reaction temperature by mixing it with the hot combustion products of fuel gas and air. This combustion may be carried out with a deficiency of air if the tail gas does not contain sufficient $H_2$ and CO to reduce all of the $SO_2$, COS, $CS_2$ and S (vapor form) to $H_2S$. The heated gas mixture is then passed through a catalyst bed (hydrogenation reactor) where all sulfur compounds are converted to $H_2S$ by hydrogenation and hydrolysis. The hydrogenated gas stream is then indirectly cooled in the reactor effluent cooler, generating steam. The gas is further cooled by direct contact with a slightly alkaline buffer solution before it enters the $H_2S$-removal portion of the process. The Stretford process is used next to remove $H_2S$ from the cooled hydrogenated tail gas, where the $H_2S$ is absorbed in an oxidizing alkaline solution. The oxidizing agents in the solution convert the $H_2S$ to elemental sulfur. The oxidizing agents are then regenerated by contacting with air in the oxidizer tank where the sulfur is floated off as a slurry. This sulfur slurry is separated from the oxidizing alkaline chemicals by filtering or centrifuging. It is reslurried with wash water and heated to melt the sulfur. The molten sulfur flows from the decanter to the sulfur pit. The oxidizing alkaline chemicals are returned to the system and the wash water is discarded. Tail gas from the absorber does not require incineration and can be vented directly to the atmosphere. An incinerator is installed as a stand-by unit and used during the start-up and shutdown of the BSRP tail gas cleanup unit and those occasions when the $H_2S$ content exceeds 10 ppmv. A flow diagram for the Beavon Sulfur Removal Process is presented in Figure 4-9.

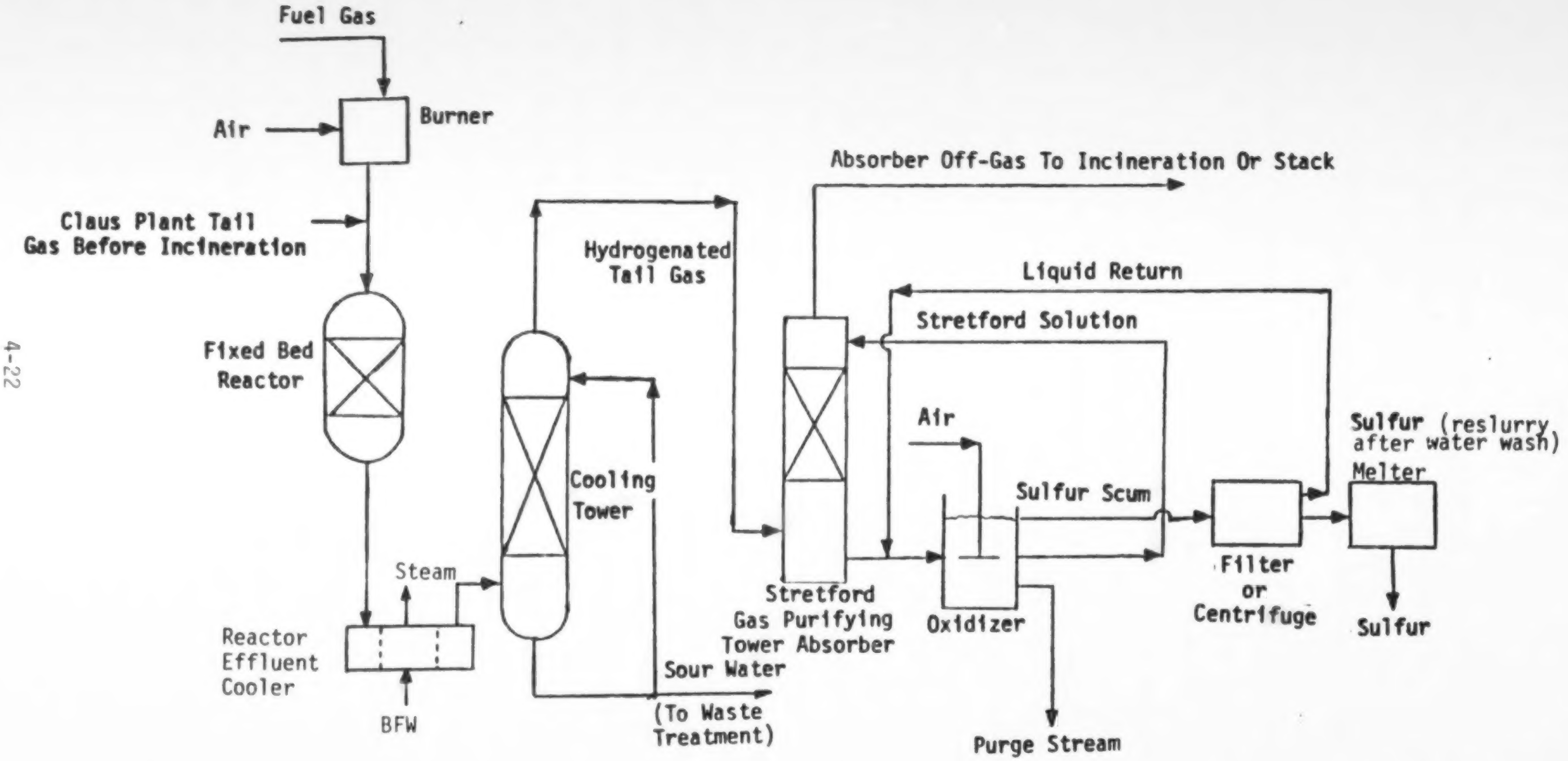
Fuel Gas
Air
Burner
Claus Plant Tail
Gas Before Incineration
Fixed Bed
Reactor
Steam
Reactor
Effluent
Cooler
BFW
Cooling
Tower
Hydrogenated
Tail Gas
Sour Water
(To Waste
Treatment)
Absorber Off-Gas To Incineration Or Stack
Stretford
Gas Purifying
Tower Absorber
Liquid Return
Stretford Solution
Air
Oxidizer
Sulfur Scum
Purge Stream
Filter
or
Centrifuge
Sulfur (reslurry
after water wash)
Melter
Sulfur

Figure 4-9. Flow diagram for the Beavon Sulfur Removal Process (BSRP).

Currently, over thirty-six BSRP plants are either operating, under construction, or in design at twenty-two locations in the United States and Japan.

4.2.5.4 3-Stage Claus Unit With BSR/Selectox I. The BSR/Selectox I process, recently developed by Union Oil Company of California and the Ralph M. Parsons Company, is a Claus tail gas cleanup system designed to provide an overall sulfur recovery efficiency in the range of 97.84 percent to 98.49 percent (average-of-run). This new process has proven to be reliable, easy to operate, and the least costly of the known tail gas cleanup processes.[4] The first industrial BSR/Selectox I unit began operation during 1978 in Lingen, West Germany.

BSR/Selectox I is a fixed bed catalytic process consisting of two steps. The first step is also part of the older Beavon Sulfur Removal Process. First, tail gas from the second stage of the Claus plant is heated to a reaction temperature of 288°C to 400°C (550°F to 750°F) in the reducing gas generator by mixing it directly with hot products from the combustion of fuel gas and air. Some hydrogen and carbon monoxide are formed to supplement the hydrogen in the tail gas. The hot gas mixture is passed through a single catalyst bed (hydrogenation reactor). The tail gas is hydrogenated to convert $SO_2$ and sulfur vapor(s) to $H_2S$ and hydrolyzed to convert COS and $CS_2$ to $H_2S$. Then, the hydrogen sulfide-containing gas stream is cooled in a contact condenser to reduce water partial pressure. The purpose of cooling is to remove water vapor, which increases conversion since water is one of the products of reaction. In the second step of the process, the cooled gas stream is reheated to a moderate temperature and mixed with a stoichiometric amount of air and passed over the Selectox-32 catalyst (proprietary catalyst of Union Oil Company) to oxidize selectively the hydrogen sulfide to elemental sulfur. Elemental sulfur is removed by condensation and is collected in a sulfur pit. At this point, total sulfur recovery is greater than 97 percent. Finally, the resulting tail gas from the BSR/Selectox I unit is passed through an additional Claus stage before incineration to increase the total sulfur recovery to more than 98 percent. Alternatively, the tail gas from the Selectox unit after sulfur condensation is routed to a thermal oxidizer and stack. A flow diagram for the BSR/Selectox I process is presented in Figure 4-10.

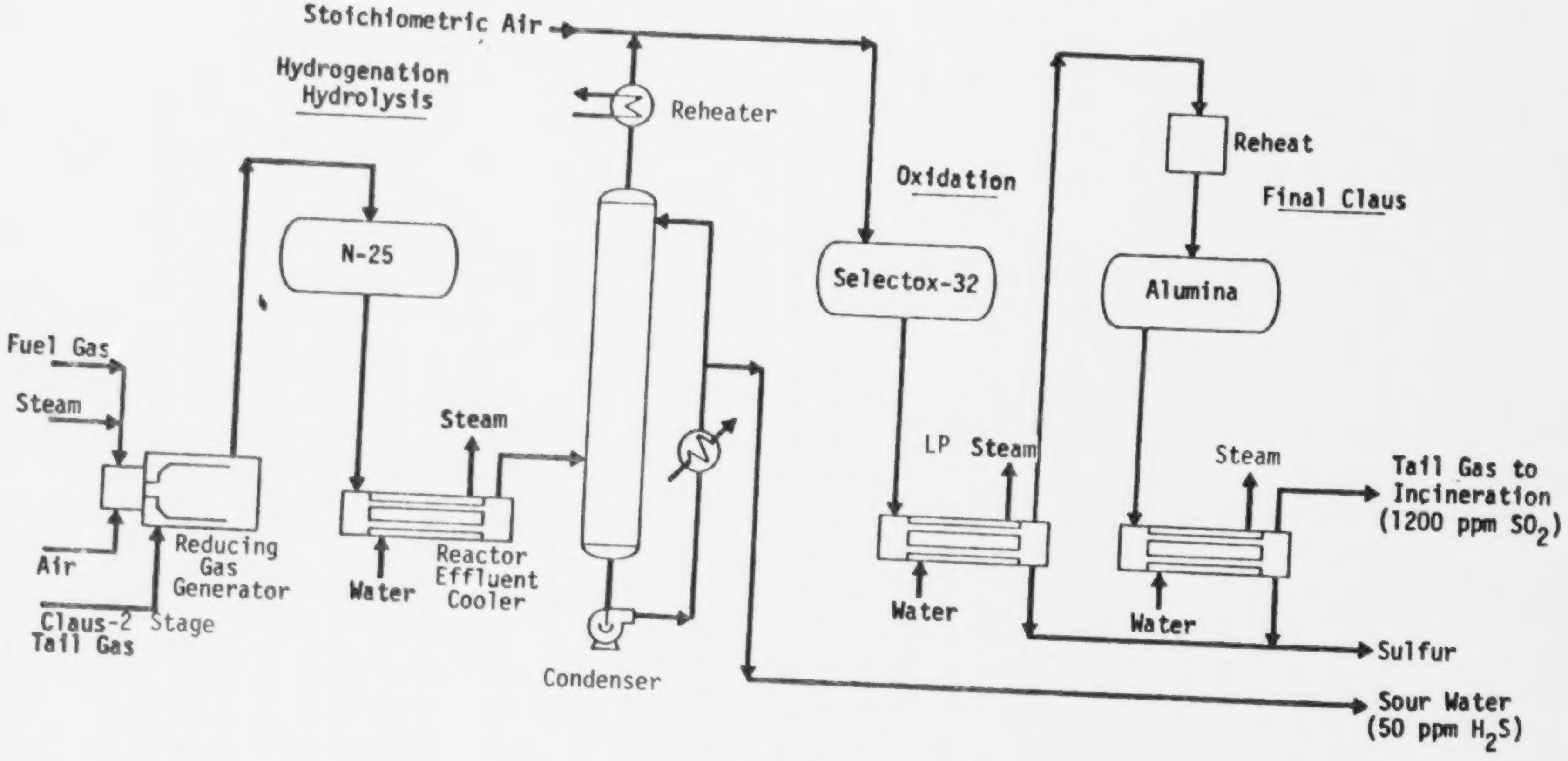
Stoichiometric Air
Hydrogenation Hydrolysis
Reheater
Oxidation
Reheat
Final Claus
N-25
Selectox-32
Alumina
Fuel Gas
Steam
Air
Claus-2 Stage Tail Gas
Reducing Gas Generator
Steam
Water
Reactor Effluent Cooler
Condenser
LP Steam
Water
Steam
Water
Tail Gas to Incineration (1200 ppm $SO_2$)
Sulfur
Sour Water (50 ppm $H_2S$)

Figure 4-10. Flow diagram for the BSR/Selectox I Claus tail gas cleanup process.

The BSR/Selectox I process recovers 80 to 90 percent of the sulfur contained in Claus plant tail gas in a continuous operation that requires no solvents or chemicals. Performance of the first industrial BSR/Selectox I unit has proved that the operation is stable and reliable. In addition, the process is very cost-effective for an overall sulfur recovery of 97.84 to 98.49 percent.[4]

4.2.5.5 Other Tail Gas Cleanup Processes. The tail gas cleanup processes described here are not widely utilized on a commercial scale in this country in the onshore natural gas production activities and are not likely to be employed in the near future. Therefore, these processes and their economics have not been analyzed in detail. A few of these processes are employed on a commercial scale elsewhere in the world.

Claus 3-stage unit with cold bed adsorption (CBA) process, developed by Amoco Production Company, is basically an extension of the Claus reaction on cold Claus catalyst bed at 127°C to 150°C (260°F to 300°F). Overall sulfur recoveries can be increased to 98 percent.[3] A final catalytic converter is provided at a low temperature to shift the reaction equilibrium to increase conversion to sulfur. Sulfur formed during this process is adsorbed on the catalyst instead of leaving the converter in the effluent vapor. Sufficient catalyst is provided to maintain the desired rate of conversion to sulfur. The catalyst is regenerated periodically to remove adsorbed sulfur and insure continued catalyst activity. When regeneration is complete, the catalyst is cooled before being used again as the CBA adsorption converter. Currently, three units (one in the United States) are in operation at Claus plants with capacities ranging from 15 to 900 metric tons of sulfur intake per day.[3]

IFP-1 (Institut Francais du Pétrole), or IFP Clauspol 1500, process developed in France is simple in design and operation, capable of achieving an overall sulfur recovery efficiency of approximately 99.3 percent.[3,9] The tail gases from a one-, two-, or three-stage Claus unit enters a vertical, packed-tower at about 127°C (260°F). Here the tail gas is contacted with a countercurrent recirculating stream of polyethylene glycol (molecular weight 400) containing a dissolved metal-salt catalyst. $H_2S$ and $SO_2$ contained in the tail gas transfers from the gas phase into liquid phase where the Claus reaction takes place. The liquid sulfur produced (99.9 percent pure) separates from the solvent and settles to the bottom of the reactor vessel, where it is removed through a seal leg. The IFP-1 reactor does not reduce COS and $CS_2$ present in the Claus tail gas feed, and, therefore, these reduced sulfur compounds remain unchanged in the reactor exhaust gas. It is essential that the $H_2S/SO_2$ ratio is kept as near to 2:1 as possible. The treated tail gas is incinerated and discharged to the atmosphere as $SO_2$. Currently, approximately twenty-five units are in operation worldwide for Claus plants ranging in size from 30 metric tons to 810 metric tons of sulfur intake per day.[3] Only a few units (non-gas applications) exist in this country.

The Cleanair process, jointly developed by the J. F. Pritchard and Company and Texas Gulf Incorporated, is designed around three separate process steps. During the first step, COS and $CS_2$ compounds are reduced by operating the Claus reactors at elevated temperatures. In the second step, the Claus tail gas is quenched to remove water and entrained sulfur and to reduce the temperature to 50°C (120°F). Then the cooled gas is fed to a reactor where $H_2S$ and $SO_2$ (in a 2:1 ratio) react, lowering the $SO_2$ concentration to less than 250 ppmv. Water and sulfur produced during this reaction are removed. The remaining $H_2S$ is oxidized to elemental sulfur in a Stretford unit which is the third step. The purified gas is then incinerated to $SO_2$ and discharged. The process is capable of recovering 99.9 percent of the sulfur from the Claus plant tail gas, leaving no more than 50 ppmv $SO_2$ equivalent in the effluent.[3] The first commercial installation was made at the Gulf Oil Corporation refinery in Santa Fe Springs, California. No unit is in operation in the onshore production industry in this country.

In IFP-2 process, licensed by Institut Francais du Pétrole of France, the Claus plant tail gas is first catalytically incinerated to oxidize all sulfur compounds to $SO_2$. The incinerated gas is cooled and then fed to an ammonia scrubber, where $SO_2$ is absorbed and converted to ammonium sulfite ($(NH_4)_2SO_3$) and ammonium bisulfite ($NH_4HSO_3$). Ammonium sulfate and thiosulfates are also formed. Gas leaving the reactor is reheated and vented to the atmosphere at less than 250 ppmv $SO_2$. The $SO_2$-rich solution is fed to an $SO_2$ regenerator, where the sulfite and bisulfite are thermally decomposed to $SO_2$, $NH_3$ and $H_2O$. Ammonium sulfate and thiosulfate in the saturated solution drawn from the bottom of the $SO_2$ regenerator are thermally decomposed in a sulfate reducer. Gases from the $SO_2$ regenerator and sulfate reducer are combined with an $H_2S$-rich stream and fed to a catalytic reactor where they are contacted with a polyethylene glycol solvent. The $H_2S$ and $SO_2$ react in the solution to form elemental sulfur. Gases from the reactor are cooled to condense out $H_2O$ and $NH_2$ as $NH_4OH$. The $NH_4OH$ solution is returned to the ammonia scrubber. Any $H_2S$ or $SO_2$ which leaves the reactor and is not absorbed by the $NH_4OH$ solution is recycled to the incinerator and from there to the ammonia scrubber. This process, when applied to Claus unit tail gas, is capable of reducing the $SO_2$ concentration in the residual tail gas to 300 ppmv or less.[11]

The Wellman-Lord process, licensed by Davy Powergas, oxidizes all the sulfur compounds present in Claus plant tail gas to $SO_2$. Next, the hot gases are cooled in a waste heat boiler, then quenched and fed to the $SO_2$ absorber. The absorber is fed a lean solution of sodium sulfite, which absorbs the $SO_2$ by reacting with it to form sodium bisulfite. The clean gases pass to the stack, while the rich bisulfite solution is regenerated to recover the sulfite solution. The $SO_2$ generated is piped back to the Claus plant feed or to other processing. Effluent levels of less than 100 ppmv $SO_2$ in the residual tail gas have consistently been achieved in commercial installations.[3] Currently, seven Wellman-Lord $SO_2$ recovery units are in operation treating Claus sulfur recovery plant tail gas. One unit was installed several years ago at the Chevron, El Segundo refinery in the United States. This process is relatively expensive.[9]

Ammonium thiosulfate process, licensed by the J. F. Prichard and Company for Coastal States Gas Corporation, recovers the sulfur contained in Claus plant tail gas as an aqueous solution of ammonium thiosulfate. The sulfur compounds in the Claus tail gas are oxidized to $SO_2$ and absorbed by contact with a weak solution of ammonia, that produces a solution of mixed ammonium bisulfite, ammonium sulfite and ammonium sulfate salts. Finally, this solution is converted to ammonium thiosulfate in a reactor. The clean tail gas contains $SO_2$ concentrations of less than 900 ppmv.[3] One unit is in operation in the United States in the onshore production field that processes 20,000 megagrams/yr sulfur intake.

### 4.2.6 Deactivation of Catalyst Activity and Reduction in the Sulfur Recovery Efficiency

The Claus sulfur recovery process involves a gas phase chemical reaction in which 2 moles of $H_2S$ combine with 1 mole of $SO_2$ to produce 3 moles of elemental sulfur. The reaction is conducted in the presence of an alumina catalyst, which increases the rate of the reaction, thereby increasing the recovery of liquid sulfur. The activity of the alumina catalyst in converting $H_2S$ and $SO_2$ to elemental sulfur depends upon its surface area: the larger the area available, the higher the reaction rate.

There are three major causes that contribute to a reduction in the catalyst surface area and catalyst activity, and thus a reduction in sulfur recovery. These are summarized as follows:

(1) sulfation of the alumina catalyst;
(2) carbon deposition on the catalyst surface due to the presence of hydrocarbons; and
(3) physical attrition of the catalyst.

Aluminum oxide reacts with the $SO_2$ and $O_2$ present in the gas stream; is converted into aluminum sulfate; and results in a gradual reduction of the available active surface area. The hydrocarbons (e.g., methane and ethane) that may be carried over from the preceding sweetening operation, in the presence of insufficient air, produce carbon particles that deposit on the catalyst surfaces and decrease the available area. Physical attrition of the catalyst in the reactor beds gradually reduce the catalyst surface area. The combined effect of aluminum sulfate and carbon deposition, as well as attrition, is to deactivate the catalyst in a period of 4 years to the point where catalyst replacement becomes necessary.

Sulfur recovery efficiency is higher when the catalyst is fresh, i.e., with an entirely active surface, and gradually declines with time to its lowest point when the catalyst is spent, i.e., deactivated to the point where it is about to be replaced. The reduction in recovery efficiency is not linear with time, however; it is significantly reduced over a period of 3 to 5 years. The sulfur recovery design study[4] indicates that the average rate of reduction in the sulfur recovery efficiency varies with the type of sulfur recovery process. The study shows that the efficiency of a Claus 2-stage when the catalyst is fresh (start-of-run efficiency) declines at an average of 1.14% every year. The efficiency when the catalysts are replaced at the end of 4 years (end-of-run efficiency) are therefore approximately 4.6% less than the start-of-run efficiency. The rates of reduction in the sulfur recovery efficiency for Recycle Selectox process were calculated from the data presented in Appendix H.

For all technologies, 4 year average catalyst life was assumed in estimating the annual average rate of reduction in sulfur recovery efficiency. For the Claus catalyst the 4 year life was based on an average of actual operating data for about 8 facilities.[15] The 4 year life assumption for the Selectox catalyst is likely conservative. Actual Selectox catalyst life probably will exceed the average Claus catalyst life because the contaminant light hydrocarbons that contribute to Claus catalyst degeneration have a less detrimental effect on the Selectox catalyst.[13] However, because only few actual data are available on the life of the Selectox catalyst, a 4 year life was assumed in estimating the average rate. The rates of reduction in the sulfur recovery efficiency for various technologies are as follows:

| Sulfur recovery technology | Average rate of reduction in sulfur recovery efficiency[4] (4 year average catalyst life) |
|---|---|
| Claus 2-stage | 1.14%/year |
| Claus 3-stage | 0.89%/year |
| Sulfreen | 0.29%/year |
| SCOT (or BSRP) | 0.013%/year |
| Recycle Selectox 2-stage | 1.68%/year |
| Recycle Selectox 3-stage | 1.88%/year |

## 4.3 COMPARISON OF SOURCE EMISSION DATA AND RALPH M. PARSONS DESIGN STUDY[4]

Table 4-1 presents a comparison of source emission test data for ten different Claus sulfur recovery facilities with the results from the Ralph M. Parsons design study.[4] These Claus facilities vary in sulfur intake size, number of Claus reactor stages utilized, type of tail gas cleanup system employed, and the $H_2S$ concentrations in the acid gas feed to the Claus sulfur recovery unit. The design sulfur recovery efficiency figures from the Parsons study are in close agreement with the source emission tests data. The data indicate a range of sulfur recovery efficiencies expected for Claus recovery facilities in relation to the $H_2S$ concentration in the acid gas feed stream and the Claus unit configuration utilized. An alphabetical designation is given for each source of Claus sulfur recovery efficiency data, "o" for operating (emission test) data and "p" for Parsons design study data.

# Table 4-1. COMPARISON OF SOURCE EMISSION TESTS DATA AND THE RALPH M. PARSONS DESIGN STUDY FOR CLAUS SULFUR RECOVERY FACILITIES

| Plant | Process equipment | Sour gas volumetric flow rate (Mm$^3$/day)* | Acid gas volumetric flow rate (Mm$^3$/day)* | Average $H_2S$ concentration in dry acid gas (volume percentage) | Liquid sulfur recovery (Mg/day) | Sulfur recovery efficiency (percent) |
|---|---|---|---|---|---|---|
| Warren Petroleum's Monument Facility[a] | 3-stage Claus unit | 2.52 (d)<br>1.72 (o) | 0.06 (o) | 24 | 32.5 (d)<br>17.8 (o) | 94.8 (o)<br>94.9 (p) |
| Getty Oil's New Hope Facility[a] | 2-stage Claus (two trains in parallel) with common 3rd stage | 1.70 (d)<br>0.77 (o) | 0.17 (o) | 55 | 152.4 (d)<br>128.0 (o) | 95.0 (o)<br>96.2 (p) |
| Shell Oil's Thomasville Facility[a] | 3-stage Claus unit | 2.83 (d)<br>2.46 (o) | 1.29 (d)<br>1.03 (o) | 84.4 | 1,295.2 (d)<br>1,174.0 (o) | 96.8 (o)<br>96.64 (p) |
| Exxon's Blackjack Creek Facility[b] | 3-stage Claus with SCOT tail gas cleanup unit | 0.85 (o) | 0.10 (o) | 86.4 | 101.4 (o) | 99.86 (o)<br>99.98 (p) |
| Shell Oil's Bryans Mill Facility[b] | 3-stage Claus unit | 1.87 (d)<br>1.83 (o) | 0.25 (d)<br>0.21 (o) | 68.9 | 253.9 (d)<br>199.0 (o) | 96.43 (o)<br>96.5 (p) |
| Shell Oil's Person Plant Facility[b] | 2-stage Claus (two trains in parallel) with common 3rd stage | 0.96 (d)<br>0.97 (o) | 0.05 (o) | 20.6 | 19.4 (o) | 95.24 (o) |
| Exxon's Santa Rosa Facility[b] | 3-stage Claus unit | 0.73 (o) | 0.10 (o) | 80.1 | 104.6 (o) | 96.5 (o)<br>96.55 (p) |
| Exxon's Flomaton Facility[b] | 3-stage Claus unit | 0.88 (d)<br>0.99 (o) | 0.52 (d)<br>0.53 (o) | 20.6 | 136.1 (d)<br>130.1 (o) | 96.7 (o)<br>94.5 (p) |
| Aquitaine's Rag River Facility[b] | 2-stage Claus unit (four Claus trains in parallel) with Sulfreen tail gas cleanup | N/A | 3.62 (o) | 84 | 3,834.0 (o) | 98.0 (o)<br>98.8 (p) |
| Chevron's Fox Creek Facility[b] | 4-stage Claus unit (two trains) | N/A | 3.40 (o) | 77 | 3,598.0 (o) | 98.65 (o |

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*At 289 K (60°F) and 1.01325 x $10^5$Pa (1.0 standard atmosphere).

[a] Detailed emission source test data developed by U.S. Environmental Protection Agency, Emissions Measurement Branch. For test methods and operating conditions, refer to Appendix C of this document. Tests conducted by Radian Corporation, Austin, Texas.

[b] Emission source test data supplied by the plant facility. For further details on the source test data supplied by the individual facility, refer to Appendix C of this document. The data from the facility's files do not specify test methods.

(d) Design values.

(o) Operating values for normal or average conditions.

(p) Derived from the Ralph M. Parsons design study. Average-of-run efficiency. Refer to Appendix E.

N/A Information not available.

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3. Gas Processing Handbook. Hydrocarbon Processing. pp. 99-170. April 1979. Docket No. A-80-20-A, Entry II-I-42.

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15. Docket No. A-80-20-A, Entry II-B-28, Average catalyst life in an onshore natural gas sulfur recovery facility. A list of eight plant facilities (that were visited) with their individual data on average catalyst life.

# 5. MODIFICATION AND RECONSTRUCTION

## 5.1 BACKGROUND

An existing facility as defined in Section 60.2 of the General Provisions (40 CFR 60)[1] is, with reference to a stationary source, an apparatus for which a standard of performance is promulgated, and the construction or modification of which commenced prior to the date of proposal of that standard. An affected facility means, with reference to a stationary source, any apparatus to which a standard is applicable. An existing facility may become an affected facility and therefore subject to the standards, through modification or reconstruction, as described below. The affected facility in the natural gas production industry is the sour natural gas sweetening operation – removal of hydrogen sulfide ($H_2S$) and carbon dioxide ($CO_2$) from sour natural gas – followed by either incineration of the acid gas stream or a sulfur recovery operation (elemental sulfur recovered from $H_2S$ in acid gas) with final incineration of unconverted $H_2S$. Emission standards promulgated under Section 111(b) of the Clean Air Act apply to all facilities within the natural gas production industry that are newly constructed, modified, or reconstructed after the date of proposal of the standard. Uncertainties may arise as to the determination of whether any existing facility has been "modified" or "reconstructed". These issues are addressed in §60.14 and §60.15[1], respectively of Title 40 of the Code of Federal Regulations Part 60, which define conditions under which an existing facility that is altered may be considered to be modified or reconstructed.

### 5.1.1 Modification

Any physical or operational change to an existing facility that results in an increase in the emission rate to the atmosphere of any pollutant to which a standard applies may be considered a modification.

Upon modification, an existing facility becomes an affected facility for each pollutant to which a standard applies and for which there is an increase in the emission rate to the atmosphere. This definition is described in §60.14. Certain physical or operational changes that are not considered a modification include:

(a) Maintenance, repair, and replacement, determined to be routine for the facility.

(b) An increase in production rate of an existing facility without a capital expenditure as defined in §60.2.

(c) An increase in the hours of operation.

(d) Use of an alternate fuel or raw material if prior to the standard, the existing facility was designed to accommodate that alternative use. A facility shall be considered to be designed to accommodate an alternate fuel or raw material if its use could be accomplished under the facility's construction specifications as amended prior to the change.

(e) The addition or use of any system or device whose primary function is the reduction of air pollutants, except when an emission control system is removed or replaced by a system that is determined less environmentally beneficial.

(f) The relocation or change in ownership of an existing facility.

The above exemptions are described in detail in §60.14.

An increase in the production rate of an existing facility is designated as a modification only if there is an increase in the emission rate and the total cost necessary to accomplish the change constitutes a "capital expenditure." Capital expenditure means an expenditure for a physical or operational change to an existing facility that exceeds the product of the applicable "annual asset guideline repair allowance percentage (AAGRAP)" specified in the latest edition of Internal Revenue Service (IRS) Publication 534 and the existing facility's original cost as defined by Section 1012 of the Internal Revenue Code.

### 5.1.2 Reconstruction

Reconstruction means the replacement of components of an existing facility to such an extent that (1) the fixed capital cost of the new components exceeds 50 percent of the fixed capital cost that would be required to construct a comparable entirely new facility, and (2) it is technologically and economically feasible to meet the applicable standards.

An existing facility, upon reconstruction, becomes an affected facility, irrespective of any change in emission rate. Fixed capital cost means the capital needed to provide all the depreciable components.

## 5.2 APPLICABILITY TO THE NATURAL GAS PRODUCTION INDUSTRY

### 5.2.1 Modifications to the Natural Gas Production Industry

This section describes typical situations in the industry and concludes that no modifications are anticipated that would subject the industry to compliance with the standard.

In a typical situation, a natural gas processing facility processes the natural gas throughput from several wells located in one reservoir or several reservoirs located in a given field. As this field is further developed, processing throughput will rise to the plant capacity (design capacity) where it will normally remain steady-state until the reservoirs are significantly depleted. During the period the processing throughput increases, the sulfur recovery will increase to near design capacity and remain at that level. Then, during the period of decreasing processing throughput, the recovery decreases gradually to low levels. Emissions of sulfur dioxide (SO) may increase or decrease during such processing variations. However, the recovery efficiency is not changed (acid gas $H_2S/CO_2$ ratio unchanged). Because a sulfur recovery facility is normally designed at the maximum anticipated recovery level, increasing or decreasing recovered sulfur production will not require any significant physical or operational changes in the recovery facility and will be accomplished without any capital expenditure. Therefore, this will not constitute a modification.

In another situation, an increase in $H_2S/CO_2$ volume percent ratio in the acid gas feedstream will increase recovery efficiency and therefore sulfur production rate. A decrease in the ratio in the acid gas feedstream will decrease recovery efficiency and therefore sulfur production rate. Emission rate increases when recovery efficiency decreases and vice versa. However, these changes are accomplished without a capital expenditure and, therefore, do not constitute a modification.

The acid gas feed stream to a sulfur recovery facility may increase above the originally designed capacity due to increased processing activities and/or introduction of other acid gas streams. This may necessitate alterations to accommodate the increased processing. The alterations may include: (1) changes to the Claus reaction furnace--increase in heat exchange capacity, auxiliary burner installation or replacement of refractory lining, or (2) any Claus process change--split flow, preheat, or inline burner. The alterations described may or may not reduce emission levels, but if the alterations incur no capital expenditure, as defined in CFR §60.2, the changes would not constitute a modification and the existing sweetening and sulfur recovery facility would not be subject to compliance with the standard.

### 5.2.2 Reconstructions to the Natural Gas Production Industry

Changes to existing facilities that would qualify as "reconstructions" are not anticipated in the natural gas production industry. The addition of a catalyst reactor stage to an existing Claus unit is not a "reconstruction" since the replacement costs are not expected to exceed 50 percent of the cost of an entirely new facility.

## 5.3 REFERENCES FOR CHAPTER 5

1. Code of Federal Regulations Title 40, Protection of Environment, Part 60. Published by the Office of the Federal Register.

# 6. MODEL PLANTS AND REGULATORY ALTERNATIVES

This chapter defines the model plants which have been selected to represent the natural gas production industry and the regulatory alternatives by which sulfur dioxide ($SO_2$) emissions from the facility can be regulated.

## 6.1 MODEL PLANTS/PARAMETERS

In order to have a common basis for calculating the various impacts of each regulatory alternative, the concept of model plants is utilized. Hypothetical plants are defined in terms of the process, production rate, raw materials, products, and other parameters that describe the industry. Usually several model plants are defined to include the types and size ranges of plants commonly found in the industry. Typically, the model plants reflect the latest technology and describe plants that are likely to be built as the industry expands. Dollar costs, resource requirements, environmental impacts, and other determinations are then made for each model plant. The results are applied in estimating economic and related effects and levels of pollution control attainable for selected control equipment and techniques.

A total of 21 model plants have been developed to project the environmental and economic impacts of various emission control alternatives on natural gas production facilities. Table 6-1 presents sulfur feed rate, acid gas $H_2S/CO_2$ ratio and regulatory alternatives for each of the 21 model plants. "Sulfur feed rate" means the long tons per day of sulfur (as hydrogen sulfide) in the acid gas emerging from the sweetening operation for the specified model plant facility. "Acid gas $H_2S/CO_2$ ratio" means volume percent ratio of hydrogen sulfide ($H_2S$) to carbon dioxide ($CO_2$) in the acid gas feed to the sulfur recovery unit. These model plants were developed using process data from existing facilities

## Table 6-1. NATURAL GAS PRODUCTION MODEL PLANTS, BASELINE CONTROLS AND REGULATORY ALTERNATIVES

| Model plant | Sulfur feed size, LT/D | $H_2S$ to $CO_2$ ratio | Regulatory alternative<br>I (baseline controls) | II | III | IV | V | VI |
|---|---|---|---|---|---|---|---|---|
| 1 | <0.1 | —[1] | None | None | — | — | — | — |
| 2 | 0.1 | —[1] | None | None | — | — | — | — |
| 3 | 0.2 | —[1] | None | — | — | — | — | Recycle Selectox 2-stage |
| 4 | 0.3 | —[1] | None | — | — | — | Recycle Selectox 2-stage | Recycle Selectox 2-stage |
| 5 | 0.4 | —[1] | None | — | — | — | Recycle Selectox 2-stage | Recycle Selectox 3-stage |
| 6 | 0.5 | —[1] | None | — | — | Recycle Selectox 2-stage | Recycle Selectox 3-stage | Sulfreen |
| 7 | 0.6 | —[1] | None | — | — | Recycle Selectox 2-stage | Recycle Selectox 3-stage | Sulfreen |
| 8 | 0.7 | —[1] | None | — | — | Recycle Selectox 2-stage | Recycle Selectox 3-stage | Sulfreen |
| 9 | 0.8 | —[1] | None | — | — | Recycle Selectox 2-stage | Recycle Selectox 3-stage | Sulfreen |
| 10 | 0.9 | —[1] | None | — | — | Recycle Selectox 2-stage | Recycle Selectox 3-stage | Sulfreen |
| 11 | 1.0 | —[1] | None | — | Recycle Selectox 2-stage | Recycle Selectox 2-stage | Recycle Selectox 3-stage | Sulfreen |
| 12 | 2.0 | —[1] | None | — | Recycle Selectox 2-stage | Recycle Selectox 2-stage | Recycle Selectox 3-stage | Sulfreen |
| 13 | 3.0 | —[1] | None | — | Recycle Selectox 2-stage | Recycle Selectox 3-stage | Sulfreen | Sulfreen |
| 14 | 4.0 | —[1] | None | — | Recycle Selectox 2-stage | Recycle Selectox 3-stage | Sulfreen | Sulfreen |
| 15 | 5.0 | —[1] | None | — | Recycle Selectox 2-stage | Recycle Selectox 3-stage | Sulfreen | Sulfreen |
| 16 | 10 | —[2] | Claus 2-stage | Claus 2-stage | Claus 3-stage | Claus 3-stage | Sulfreen | Sulfreen |
| 17 | 100 | —[2] | Claus 3-stage | Claus 3-stage | Sulfreen | Sulfreen | SCOT/BSRMDEA or BSRP [3] | SCOT/BSRMDEA or BSRP [3] |
| 18 | 555 | 50/50 | Claus 3-stage | Sulfreen | Sulfreen | SCOT/BSRMDEA or BSRP [3] | SCOT/BSRMDEA or BSRP [3] | SCOT/BSRMDEA or BSRP [3] |
| 19 | 555 | 80/20 | Claus 3-stage | Sulfreen | SCOT/BSRMDEA or BSRP [3] | SCOT/BSRMDEA or BSRP [3] | SCOT/BSRMDEA or BSRP [3] | SCOT/BSRMDEA or BSRP [3] |
| 20 | 1,000 | 50/50 | Claus 3-stage | Sulfreen | Sulfreen | SCOT/BSRMDEA or BSRP [3] | SCOT/BSRMDEA or BSRP [3] | SCOT/BSRMDEA or BSRP [3] |
| 21 | 1,000 | 80/20 | Claus 3-stage | Sulfreen | SCOT/BSRMDEA or BSRP [3] | SCOT/BSRMDEA or BSRP [3] | SCOT/BSRMDEA or BSRP [3] | SCOT/BSRMDEA or BSRP [3] |

[1] Covers the entire ratio range from less than 2/98 to over 12.5/87.5.

[2] Covers the entire ratio range from less than 12.5/87.5 to over 80/20.

[3] SCOT – Shell Claus Offgas Treatment
BSRMDEA – Beavon Sulfur Removal Methyldiethanolamine (MDEA), BSRMDEA equivalent to SCOT
BSRP – Beavon Sulfur Removal Process

along with technical data from a July 1982 report by the American Petroleum Institute (API) on their gas plant survey, from meeting reports, from plant visit reports, and from a review of the published studies. Appendix G contains the API survey information for model plants with feed rates of 5 LT/D or less.

### 6.1.1 Model Plant Sizes[1]

The sizes (sulfur feed rates) of the model plants specified in Table 6-1 span the sizes currently existing in the industry and are also representative of the sweetening and sulfur recovery capacity of projected future plants within the industry. The first 15 small size model plants represent sulfur feed rates of less than 0.1 Mg/d (0.1 LT/D) through 5.0 Mg/d (5 LT/D), and the largest model plant represents 1,016 Mg/d (1,000 LT/D) of sulfur intake to sulfur recovery unit. The sizes of the remaining model plants are 10.2 Mg/d (10 LT/D), 102 Mg/d (100 LT/D) and 564 Mg/d (555 LT/D).

### 6.1.2 $H_2S/CO_2$ Volume Percent Ratio

The ratio of volume percent concentration of $H_2S$ to that of $CO_2$ in the acid gas feed to the sulfur recovery unit was considered in the development of each model plant. The ratio, as found in the industry practice, ranged from less than 2 percent $H_2S$/98 percent $CO_2$ to over 80 percent $H_2S$/20 percent $CO_2$.

### 6.1.3 Baseline Control Levels

Baseline controls are defined as those levels of control expected to be utilized in new facilities in the absence of a standard. Baseline controls are used to assess incremental environmental, economic, and energy impacts of each regulatory alternative applied to the model plants.

Baseline control technology for sweetening and sulfur recovery plants was chosen according to the amount of sulfur available for processing. Generally the larger the rate of sulfur feed rate and the higher the concentration of $H_2S$ in the acid gas, the higher the degree of control of emitted $SO_2$.

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[1] Detailed parameters for the model plants with sulfur feed rates of 5 through 1,000 LT/D are summarized in Docket Entry A-80-20-A, II-B-18. The parameters included sulfur feed rate (sulfur intake), acid gas $H_2S/CO_2$ ratio, sulfur recovery efficiency and stack gas characteristics. The information was extracted from Appendix E.

Three levels of control have been determined to specify baseline control technologies in the model plants. Existing State regulations and existing control levels in current industry practice were reviewed in determining these control levels. Table 6-1 presents baseline control levels for each model plant (see Regulatory Alternative I). Model Plants 1 through 15 incorporate sweetening of sour natural gas, with thermal oxidation to $SO_2$ of $H_2S$ separated in the sweetening unit into $SO_2$ and subsequent release through an incinerator stack into the ambient atmosphere (no sulfur recovery). Model Plant 16 incorporates sweetening of sour natural gas, with Claus 2-stage sulfur recovery from acid gas, with thermal oxidation of unconverted $H_2S$ into $SO_2$ and subsequent release into the ambient atmosphere through an incinerator stack. Model Plants 17 through 21 incorporate sweetening of sour natural gas, with Claus 3-stage sulfur recovery from acid gas, with thermal oxidation of unconverted $H_2S$ into $SO_2$ and subsequent release into the ambient atmosphere through an incinerator stack.

## 6.2 REGULATORY ALTERNATIVES

The purpose of this section is to define the various regulatory alternatives and consider their effectiveness for reducing baseline control emissions. The effects of a regulatory alternative can be assessed from a summation of its effects on a combination of individual model plants. Six regulatory alternatives have been developed to perform analyses on the model plants. These regulatory alternatives are summarized in Table 6-1. These alternatives entail various degrees of emission control and are developed based upon cost effectiveness and incremental cost effectiveness values derived from design data and cost figures provided by the Ralph M. Parsons Company. This information is presented in detail in Appendices E and H. The impacts of each alternative will be evaluated in the economic, environmental, and energy analyses. Sulfur recovery control technologies presented in Table 6-1 represent those technologies currently in use in the industry and expected to be used for many years.

Under Regulatory Alternative I, no new source performance standard (NSPS) would be promulgated for the natural gas production industry. This alternative uses baseline emission controls. Regulatory

Alternatives II through VI were selected as groups of processes that display increasing values of cost effectiveness and incremental cost effectiveness. "Cost effectiveness" is defined as the difference between the annualized cost of the given regulatory alternative and that of the baseline control divided by the difference in the annual long ton $SO_2$ emission reduction of the given regulatory alternative and that of the baseline control. "Incremental cost effectiveness" is the ratio of the additional cost and the additional emission reduction for moving from one regulatory alternative to the next more stringent alternative.

# 7. ENVIRONMENTAL IMPACT

The environmental impacts associated with each regulatory alternative are discussed with respect to both primary and secondary incremental impacts on air, water, solid waste, and energy resulting from the use of alternative control systems. Impacts of establishing emission standards (based upon application of the different control systems) are compared with the impacts of not proposing or promulgating standards of performance for new sources. Both beneficial and adverse impacts are assessed for each of the model plants presented in Chapter 6. The regulatory alternatives described for the model plants consist of various sulfur recovery and/or tail gas treatment units. The applicable sulfur recovery efficiencies are applied to predict the long-term effects on nationwide emissions that could result from promulgation of each regulatory alternative.

## 7.1 AIR POLLUTION IMPACT

### 7.1.1 Dispersion Modeling Results

As part of an air pollution impact study, a dispersion modeling analysis of the regulatory alternatives for the model plants with sulfur feed rates of 5 Mg/d and greater (Table 6-1) was conducted. The locations of natural gas processing facilities are generally in non-urban areas; moreover, $SO_2$ is released into the atmosphere through a tall stack. Therefore, a single source model (CRSTER) was used. It was also determined that no treatment of (building) downwash was required. The model was used to calculate maximum concentration (ambient) for averaging periods of 1-hour, 3-hour, 24-hour and a year at various radial distances. The six radial locations ranged from 0.4 to 15 kilometers. Meteorology data from locations in West Texas and the Western Gulf Basin regions were reviewed. Two surface and upper air data sets, one for Houston and the other for Amarillo, were selected. In total, 72 scenarios were tested, twelve model plant sulfur feed rates each with six regulatory alternatives. Table 7-1 summarizes the results for the 1-hour, 3-hour, 24-hour and annual averaging periods for both the Houston data and the Amarillo data. For each model plant and regulatory alternative combination, the highest second high (for short-term periods) or highest (for the annual average) $SO_2$ concentration in microgram per cubic meter is presented, independent of location distance. The results indicate that all of the scenarios tested meet national ambient air quality standard (NAAQS) levels. The methods, data bases and results are documented in the report available in Docket Entry No. A-80-20-A, II-A-18.

### 7.1.2 Effects of Regulatory Alternatives on Nationwide $SO_2$ Emissions

The summary of impacts for the model plants with each of the six regulatory alternatives (presented in Table 6-1) was extrapolated to estimate the national air quality impact over the period from 1983 through 1987. Using past onshore natural gas sulfur recovery facility growth (Appendix G and Tables 9-8 and 9-9) that is based upon the American Petroleum Institute's July 1982 survey on the onshore natural gas processing facilities, and the data on new onshore natural gas production from the American Gas Association (AGA), the future growth of the industry for the time period of interest was projected. Historically, natural gas produced offshore has been sweet. The EPA assumed that the natural gas produced offshore during the projecting period would continue to be sweet. Therefore, gas produced offshore was not considered for the growth projections. Information outlining the methodology used to develop this 5-year projection is contained in Chapter 9.

The projected total new onshore natural gas processing capacities for 1983 through 1987 with sweetening and those with sweetening as well as sulfur recovery are listed in Table 9-21. The difference between these two yields the new production without sulfur recovery. The $H_2S$ removed from this natural gas is incinerated and released to the atmosphere. This estimates nationwide uncontrolled $SO_2$ emissions. The number of projected facilities not recovering sulfur may then be calculated from the above total uncontrolled $SO_2$ emissions. The number of uncontrolled

## Table 7-1. SUMMARY OF $SO_2$ CONCENTRATIONS FROM DISPERSION MODELING ANALYSES FOR EACH MODEL PLANT SULFUR FEED RATE AND REGULATORY ALTERNATIVE ON THE BASIS OF THE HOUSTON AND AMARILLO DATA

(On the Basis of the Houston Data)

| [illegible] | [illegible] | Regulatory Alternative^a | | | | | | | | | | | | | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | I | | | | II | | | | III | | | | IV | | | | V | | | | VI | | | |
| | | 1 Hour | 3 Hour | 24 Hour | Annual | 1 Hour | 3 Hour | 24 Hour | Annual | 1 Hour | 3 Hour | 24 Hour | Annual | 1 Hour | 3 Hour | 24 Hour | Annual | 1 Hour | 3 Hour | 24 Hour | Annual | 1 Hour | 3 Hour | 24 Hour | Annual |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] |

^a Highest second-high in micrograms per $m^3$ for 1 Hour, 3 Hour, and 24 Hour averaging periods; highest in micrograms per $m^3$ for annual averaging period.

(On the Basis of the Amarillo Data)

| [illegible] | [illegible] | Regulatory Alternative^a | | | | | | | | | | | | | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | I | | | | II | | | | III | | | | IV | | | | V | | | | VI | | | |
| | | 1 Hour | 3 Hour | 24 Hour | Annual | 1 Hour | 3 Hour | 24 Hour | Annual | 1 Hour | 3 Hour | 24 Hour | Annual | 1 Hour | 3 Hour | 24 Hour | Annual | 1 Hour | 3 Hour | 24 Hour | Annual | 1 Hour | 3 Hour | 24 Hour | Annual |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | - | - | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| [illegible] | [illegible] | - | - | - | - | - | - | - | - | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | - | - | - | - | - | - | - | - |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | - | - | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | - | - | - | - | - | - | - | - | - | - | - | - | - | - | - | - |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | - | - | - | - | - | - | - | - | - | - | - | - |
| [illegible] | [illegible] | - | - | - | - | - | - | - | - | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] |
| [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | [illegible] | - | - | - | - | - | - | - | - | - | - | - | - | - | - | - | - |

^a Highest second-high in micrograms per $m^3$ for 1 Hour, 3 Hour, and 24 Hour averaging periods; highest in micrograms per $m^3$ for annual averaging period.

facilities of sizes (sulfur feed rates) varying from less than 0.1 megagrams per day through 5.0 megagrams per day that will be constructed between 1983 and 1987 was estimated to be 37 (Table G-7, Appendix G). By considering the numbers of facilities both recovering and not recovering sulfur, the nationwide $SO_2$ emissions from each of the regulatory alternatives may be projected for 1983-1987. Table 7-2 lists a summary of sulfur recovery efficiencies for the model plant regulatory alternatives.

Data on sulfur feed rate and acid gas $H_2S/CO_2$ ratio for gas streams (available from the American Petroleum Institute's Gas Plant Survey Report presented in Appendix G) were reviewed. Percent distribution of the six acid gas ratio (2/98, 5/95, 12.5/87.5, 20/80, 50/50, and 80/20) were developed for each of the five sulfur intake range categories: <5, 10, 102, 564 and 1,016 Mg/d. The distribution reflects the percentage of the gas streams that produce a certain acid gas $H_2S/CO_2$ ratio.

For each model plant, the product of the daily sulfur feed rate and the number of projected new sweetening and sulfur recovery facilities (Table 9-23 and Table G-6, Appendix G) produces the total daily sulfur feed rate. The projected daily $SO_2$ emissions are calculated by the product of the sulfur feed rate, the fraction of the gas streams that produce a certain acid gas feed $H_2S/CO_2$ ratio, the sulfur penetration rate (1 minus sulfur recovery efficiency fraction), and the $SO_2$ conversion factor (64 Mg $SO_2$/32 Mg S = 2.0). A typical operating schedule of 350 days per year was assumed to calculate annual $SO_2$ emissions by the end of 1987. Tables 7-3 and 7-4 quantify the estimated daily and annual $SO_2$ emissions, respectively, resulting from promulgation of a standard based upon each of Regulatory Alternatives I through VI. Table 7-5 summarizes the potential $SO_2$ emissions reductions beyond baseline for each of Regulatory Alternatives II through VI. Listed are the 5-year projected emission reductions and the percent reductions beyond current control levels (Regulatory Alternative I or baseline emissions). There would be a reduction from the baseline (Alternative I) of 25.6 percent with Alternative II, 78.4 percent with Alternative III, 80.5 percent with Alternative IV, 93.3 percent with Alternative V, and 93.6 percent for Alternative VI.

Table 7-2. SUMMARY OF SULFUR RECOVERY EFFICIENCIES FOR THE MODEL PLANT REGULATORY ALTERNATIVES

| Sulfur feed rate[a], Mg/d | Number of new facilities[a] | Total sulfur feed rate[b], Mg/d | Acid gas feed $H_2S/CO_2$ ratio by by volume[c] | Percent distribution of $H_2S/CO_2$ ratio[d] | Regulatory alternative sulfur recovery efficiency percent[e] | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | I (baseline) | II | III | IV | V | VI |
| <0.1-5.0[f] | 37 | 38.9 | 2/98<br>5/95<br>12.5/87.5 | 35.5<br>51.6<br>12.9 | 0.00 | 0.00 | 0.00<br>to<br>89.01 | 0.00<br>to<br>91.36 | 0.00<br>to<br>97.50 | 0.00<br>to<br>97.50 |
| 10 | 18 | 180 | 12.5/87.5<br>20/80<br>50/50<br>80/20 | 2.9<br>34.3<br>37.1<br>25.7 | 91.30<br>91.80<br>93.80<br>95.10 | 91.30<br>91.80<br>93.80<br>95.10 | 93.39<br>94.50<br>96.10<br>96.64 | 93.39<br>94.50<br>96.10<br>96.64 | 97.50<br>97.80<br>98.42<br>98.50 | 97.50<br>97.80<br>98.42<br>98.50 |
| 102 | 9 | 918 | 12.5/87.5<br>20/80<br>50/50<br>80/20 | 0.0<br>7.7<br>30.8<br>61.5 | 93.39<br>94.50<br>96.10<br>96.64 | 93.39<br>94.50<br>96.10<br>96.64 | 97.50<br>97.80<br>98.42<br>98.50 | 97.50<br>97.80<br>98.42<br>98.50 | 99.87<br>99.90<br>99.95<br>99.98 | 99.87<br>99.90<br>99.95<br>99.98 |
| 564 | 2 | 1,128 | 50/50<br>80/20 | 0.0<br>100.0 | 96.10<br>96.64 | 98.42<br>98.50 | 98.42<br>99.98 | 99.95<br>99.98 | 99.95<br>99.98 | 99.95<br>99.98 |
| 1,016 | 1 | 1,016 | 50/50<br>80/20 | 0.0<br>100.0 | 96.10<br>96.64 | 98.42<br>98.50 | 98.42<br>99.98 | 99.95<br>99.98 | 99.95<br>99.98 | 99.95<br>99.98 |

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[a] Based on 5-year projection (1983-1987) of new sweetening and sulfur recovery facilities added. Refer to Table 9-23, Appendix G and Appendix H.

[b] Refers to combined total of the sulfur feed rate.

[c] Based on API gas plant survey data, these acid gas feed ratios were found.

[d] Based on API gas plant survey data. Refers to percent of gas streams with applicable $H_2S/CO_2$ ratio by volume.

[e] Based on sulfur recovery efficiencies (average-of-run) from the Ralph M. Parsons' sulfur recovery study report. Also refer to Table 6-2, Appendix E, Tables E-25, E-26 and E-27 and Appendix H, Table H-6.

[f] This sulfur feed rate range refers to onshore natural gas processing facilities that normally do not recover sulfur. However, under the Regulatory Alternatives III through VI certain plant sizes within this range recover sulfur; other plant sizes are uncontrolled (refer to Table 6-1).

## Table 7-3. DAILY $SO_2$ EMISSIONS FROM THE PROJECTED (1983-1987) NEW ONSHORE NATURAL GAS PROCESSING FACILITIES FOR THE MODEL PLANT REGULATORY ALTERNATIVES

| Sulfur feed rate[a], Mg/d | Number of new facilities[a] | Total sulfur feed rate[b], Mg/d | Acid gas feed $H_2S/CO_2$ ratio by by volume[c] | Percent distribution of $H_2S/CO_2$ ratio[d] | Regulatory alternative projected $SO_2$ emissions, Mg/d | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | I (baseline) | II | III | IV | V | VI |
| <0.1-5.0[e] | 37 | 38.9 | 2/98 | 35.5 | 77.8 | 77.8 | 22.3 | 15.6 | 13.0 | 12.2 |
| | | | 5/95 | 51.6 | | | | | | |
| | | | 12.5/87.5 | 12.9 | | | | | | |
| 10 | 18 | 180 | 12.5/87.5 | 2.9 | 0.9 | 0.9 | 0.7 | 0.7 | 0.3 | 0.3 |
| | | | 20/80 | 34.3 | 10.1 | 10.1 | 6.8 | 6.8 | 2.7 | 2.7 |
| | | | 50/50 | 37.1 | 8.3 | 8.3 | 5.2 | 5.2 | 2.1 | 2.1 |
| | | | 80/20 | 25.7 | 4.5 | 4.5 | 3.1 | 3.1 | 1.4 | 1.4 |
| 102 | 9 | 918 | 12.5/87.5 | 0.0 | - | - | - | - | - | - |
| | | | 20/80 | 7.7 | 7.8 | 7.8 | 3.1 | 3.1 | 0.1 | 0.1 |
| | | | 50/50 | 30.8 | 22.1 | 22.1 | 8.9 | 8.9 | 0.3 | 0.3 |
| | | | 80/20 | 61.5 | 37.9 | 37.9 | 16.9 | 16.9 | 0.2 | 0.2 |
| 564 | 2 | 1,128 | 50/50 | 0.0 | - | - | - | - | - | - |
| | | | 80/20 | 100.0 | 75.8 | 33.8 | 0.5 | 0.5 | 0.5 | 0.5 |
| 1,016 | 1 | 1,016 | 50/50 | 0.0 | - | - | - | - | - | - |
| | | | 80/20 | 100.0 | 68.3 | 30.5 | 0.4 | 0.4 | 0.4 | 0.4 |
| Total | | | | | 314 | 234 | 67.9 | 61.2 | 21.0 | 20.3 |

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[a] Based on 5-year projection (1983-1987) of new sweetening and sulfur recovery facilities added. Refer to Table 9-23 and Appendix G.

[b] Refers to combined total of the sulfur feed rate.

[c] Based on API gas plant survey data, these acid gas feed ratios were found.

[d] Based on API gas plant survey data. Refers to percent of gas streams with applicable $H_2S/CO_2$ ratio by volume.

[e] This sulfur feed rate range refers to onshore natural gas processing facilities that normally do not recover sulfur. However, under the Regulatory Alternatives III through VI certain plant sizes within this range recover sulfur; other plant sizes are uncontrolled (refer to Table 6-1).

## Table 7-4. ANNUAL $SO_2$ EMISSIONS FROM THE PROJECTED (1983-1987) NEW ONSHORE NATURAL GAS PROCESSING FACILITIES FOR THE MODEL PLANT REGULATORY ALTERNATIVES

| Sulfur feed rate[a], Mg/d | Number of new facilities[a] | Total sulfur feed rate[b], Mg/d | Acid gas feed $H_2S/CO_2$ ratio by by volume[c] | Percent distribution of $H_2S/CO_2$ ratio[d] | Regulatory alternative projected $SO_2$ emissions,[e] $10^3$ Mg/y | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | I (baseline) | II | III | IV | V | VI |
| <0.1-5.0[f] | 37 | 38.9 | 2/98<br>5/95<br>12.5/87.5 | 35.5<br>51.6<br>12.9 | 27.2 | 27.2 | 7.8 | 5.5 | 4.6 | 4.3 |
| 10 | 18 | 180 | 12.5/87.5<br>20/80<br>50/50<br>80/20 | 2.9<br>34.3<br>37.1<br>25.7 | 8.4 | 8.4 | 5.5 | 5.5 | 2.3 | 2.3 |
| 102 | 9 | 918 | 12.5/87.5<br>20/80<br>50/50<br>80/20 | 0.0<br>7.7<br>30.8<br>61.5 | 23.7 | 23.7 | 10.2 | 10.2 | 0.2 | 0.2 |
| 564 | 2 | 1,128 | 50/50<br>80/20 | 0.0<br>100.0 | 26.5 | 11.8 | 0.2 | 0.2 | 0.2 | 0.2 |
| 1,016 | 1 | 1,016 | 50/50<br>80/20 | 0.0<br>100.0 | 23.9 | 10.7 | 0.1 | 0.1 | 0.1 | 0.1 |
| Total | | | | | 110 | 81.8 | 23.8 | 21.4 | 7.4 | 7.0 |

[a] Based on 5-year projection (1983-1987) of new sweetening and sulfur recovery facilities added. Refer to Table 9-23 and Appendix G.

[b] Refers to combined total of the sulfur feed rate.

[c] Based on API gas plant survey data, these acid gas feed ratios were found.

[d] Based on API gas plant survey data. Refers to percent of gas streams with applicable $H_2S/CO_2$ ratio by volume.

[e] Based on 350 days/year operating schedule and represents projected fifth-year annual $SO_2$ emissions for each size category in 1987.

[f] This sulfur feed rate range refers to onshore natural gas processing facilities that normally do not recover sulfur. However, under the Regulatory Alternatives III through VI certain plant sizes within this range recover sulfur; other plant sizes are uncontrolled (refer to Table 6-1).

Table 7-5. EFFECTIVENESS OF THE REGULATORY ALTERNATIVES IN REDUCING SO$_2$[a] EMISSIONS FROM THE PROJECTED NEW ONSHORE NATURAL GAS PROCESSING FACILITIES[a]

| | Regulatory Alternative II | Regulatory Alternative III | Regulatory Alternative IV | Regulatory Alternative V | Regulatory Alternative VI |
|---|---|---|---|---|---|
| SO$_2$ Emission Reductions[a] ($10^3$ Mg/y) | 28.2 | 86.2 | 88.6 | 102.6 | 103.0 |
| SO$_2$ Emission Reductions[a] (percent) | 25.6 | 78.4 | 80.5 | 93.3 | 93.6 |

[a] Represents total reductions beyond Regulatory Alternative I (baseline level) for the fifth year (1987).

### 7.1.3 Secondary Impacts on Air Quality

Secondary air pollutants are those emissions that are not usually associated with an uncontrolled facility but that result from the use of pollution control equipment (i.e. the control of one pollutant results in the production of another pollutant). The use of sulfur recovery units and the type of emission control systems considered in each of the regulatory alternatives would not result in any adverse secondary impacts.

## 7.2 WATER POLLUTION IMPACT

There would be essentially no water pollution impact associated with any of the six regulatory alternatives. The by-products thiosulfate ($Na_2S_2O_3$) and sulfite ($Na_2SO_3$) do tend to build up in solutions during the Stretford process and this build-up reduces the ability of the system to remove hydrogen sulfide ($H_2S$). Formation of thiosulfate can be controlled , however, to below 1% of the $H_2S$ by proper plant operation.[2,3] The Stretford process is used in the Beavon Sulfur Removal Process (BSRP) tail gas cleanup system (described in Chapter 4). The BSRP tail gas cleanup system is specified as one of two possible tail gas cleanup systems in Regulatory Alternatives III through VI (Table 6-2). The by-product thiosulfate and sulfite is transferred by a purge stream for further salt recovery treatment. Salt recovery methods that are currently available include evaporation or spray drying, biological degradation, and oxidative combustion. After salt recovery, the solid waste is buried in sanitary landfills. Another salt recovery method currently available is reductive incineration. Solids that were recovered from spray evaporation or oxidative combustion can be reduced to a product that can be recycled through the Stretford process resulting in economic savings and zero effluent discharge.[4] Therefore, liquid waste disposal problems are not considered significant.

The other sulfur recovery processes and tail gas cleanup systems (presented in Chapter 4) specified as regulatory alternatives produce no significant adverse water quality impacts, as the liquid sulfur is recovered for sale.

## 7.3 SOLID WASTE DISPOSAL IMPACT

There would be no significant adverse impacts on the level of solid waste produced as a result of promulgation of any one of the six Regulatory Alternatives specified in Table 6-1. Instead, there is an economic incentive to recover the sulfur (which is recovered in liquid, not solid form) for sale to the sulfuric acid manufacturing industry. Small plants with sulfur feed rates below 5 Mg/d may not always be able to market and sell the recovered sulfur, because the recovered amount would be small, or the market may not be accessible. These plants normally dispose of the recoverd sulfur in a landfill under such circumstances. However, the disposal impact is not expected to be significant. The Stretford process produces by-products (thiosulfate and sulfite).[2,3] These chemicals can be disposed of or recovered, creating an insignificant solid waste impact. The other possible solid wastes that are typically buried in a sanitary landfill include: 1) spent catalysts from the reactor beds (catalyst replacement depends on the replacement schedule, normally every 3 to 5 years); 2) recovered sulfur resulting from spillage or leakage when it is transferred either to the liquid sulfur storage tank or transferred for disposal or sale; and 3) spent carbon absorbers from the hydrocarbon (HC) recovery unit (carbon absorbers typically last several years). The spent catalyst is often used as road gravel (paving material) and has no adverse environmental effect.

## 7.4 ENERGY IMPACTS

The utility requirements associated with Regulatory Alternative I (the baseline requirements) for all the projected new onshore natural gas processing facilities are presented in Table 7-6. Fifth-year (1987) requirements (beyond the baseline requirements) of electric power, fuel gas, 600 psig steam, 50 psig steam, treated boiler feed water and cooling water for the projected new facilities for all the regulatory alternatives are presented in Table 7-7. The utility requirements were based on the Ralph M. Parsons Company's study of onshore natural gas processing facilities.[5] The data on the number of new onshore natural gas processing facilities and the representative sizes (sulfur feed rates in megagrams per day) of these projected facilities are presented in Table 9-23 and Tables G-6 and G-7, Appendix G. Nationwide fifth-year (1987) utility

Table 7-6. FIFTH-YEAR (1987) BASELINE (REGULATORY ALTERNATIVE I) ENERGY REQUIREMENTS FOR EACH OF THE PROJECTED NEW ONSHORE NATURAL GAS PROCESSING FACILITIES[a]

| Model plant # | Sulfur feed rate Mg/d | Number new facilities | Acid gas feed $H_2S/CO_2$ ratio by volume | Electric power, kW | Fuel gas, $10^9$ J/h | 600 PSIG steam, kg/h | 50 PSIG steam, kg/h | Treated boiler feed wtaer, kg/h | Cooling water, $m^3/h$ |
|---|---|---|---|---|---|---|---|---|---|
| 1-15 | <0.1-5 | 37 | b | 0 | 0 | 0 | 0 | 0 | 0 |
| 16 | 10 | 18 | c | 59 | 3.9 | 591 | 0 | 1,400 | 0 |
| 17 | 102 | 9 | c | 308 | 40.0 | 0 | 0 | 29,600 | 0 |
| 18 | 564 | 0 | 50/50 | 915 | 72.9 | 0 | 0 | 91,900 | 0 |
| 19 | 564 | 2 | 80/20 | 817 | 48.7 | 0 | 0 | 83,600 | 0 |
| 20 | 1,016 | 0 | 50/50 | 1,880 | 131 | 0 | 0 | 166,000 | 0 |
| 21 | 1,016 | 1 | 80/20 | 1,680 | 87.7 | 0 | 0 | 151,000 | 0 |

[a] Refer to Table 9-23 and Appendix G.

[b] Acid gas $H_2S/CO_2$ volume percent ratio range of less than 2/98 to over 12.5/87.5 considered.

[c] Acid gas $H_2S/CO_2$ volume percent ratio of 12.5/87.5 considered as worst case.

### Table 7-7. FIFTH-YEAR (1987) ENERGY REQUIREMENTS BEYOND THE BASELINE (REGULATORY ALTERNATIVE I) FOR ALL OF THE PROJECTED NEW ONSHORE NATURAL GAS PROCESSING FACILITIES[a] FOR THE REGULATORY ALTERNATIVES[b]

| | Regulatory alternative | | | | | |
|---|---|---|---|---|---|---|
| | I | II | III | IV | V | VI |
| Electric power, $10^6$ kWh/y | 0.0 | 13.2 | 62.8 | 65.1 | 102 | 100 |
| Fuel gas, $10^{14}$ J/y | 0.0 | 0.0 | 5.57 | 6.26 | 15.6 | 15.8 |
| 600 PSIG steam, $10^7$ kg/y | 0.0 | 0.0 | 5.6 | 7.3 | (4.1) | (7.3) |
| 50 PSIG steam, $10^8$ kg/y | 0.0 | 0.0 | 3.0 | 3.0 | 29.6 | 29.6 |
| Treated boiler feed water, $10^8$ kg/y | 0.0 | (0.7) | (7.2) | (7.6) | (9.1) | (7.7) |
| Cooling water, $10^7$ $m^3$/y | 0.0 | 0.0 | 2.3 | 4.0 | 10.9 | 10.9 |

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[a] Refer to Table 9-23 and Tables G-6 and G-7 in Appendix G.

[b] Refer to Table 6-1.

requirements were developed assuming 350 days per year, 24 hours per day facility operation. There would be an increase in energy (utilities) usage over Regulatory Alternative I of 0.88 percent with Alternative II; 14.5 percent with Alternative III; 15.9 percent with Alternative IV; 35.8 percent with Alternative V; and 36.03 percent with Alternative VI. These increased energy requirements result from greater sulfur recovery efficiency requirements for small, medium, and large size sweetening and sulfur recovery facilities.

## 7.5 OTHER ENVIRONMENTAL CONCERNS

### 7.5.1 Irreversible and Irretrievable Commitment of Resources

The six regulatory alternatives defined in Chapter 6 would not preclude the development of future control options nor would they curtail any beneficial use of environmental resources. No long-term environmental losses would result from the regulatory alternatives.

### 7.5.2 Environmental Impact of Delayed Standards

The only environmental impact on air pollution associated with a delay in proposing and promulgating the standard would be an increase in $SO_2$ emissions attributable to the construction of new sulfur recovery units without tail gas treatment systems. Additional $SO_2$ emissions from the currently uncontrolled small facilities with sulfur intakes less than 10 megagrams per day would result provided that a standard based on Regulatory Alternatives III, IV, V, or VI is delayed.

Delaying the standard would result in possible water and solid waste impact reductions, but the related reductions would be minimal compared with the air quality benefits attributable to promulgation of the standard.

Energy utilization by the industry would be less if the standard were delayed; however, the industry can reduce this additional expense with credits derived from the sale of the recovered sulfur.

## 7.6 REFERENCES FOR CHAPTER 7

1. "State Air Laws." Environmental Reporter. The Bureau of National Affairs, Inc. Washington, D.C. 1979-1981. Docket No. A-80-20-A, II-I-65.

2. A.J. Moyes and J.S. Wilkinson. "High-Efficiency Removal of $H_2S$ from Fuel Gases and Process Gas Streams." Process Engineering, September 1973, pp. 101-105. Docket No. A-80-20-A, Entry II-I-72.

3. A.L. Kohl and F.C. Riesenfeld. Gas Purification, 3rd ed. Gulf Publishing Company. Houston, Texas. 1979., pp. 476-487. Docket No. A-80-20-A, Entry II-I-73.

4. Srini Vasan. "Holmes-Stretford Process Offers Economic $H_2S$ Removal." Oil and Gas Journal. January 2, 1978. Docket No. A-80-20-A, Entry II-I-31.

5. Ralph M. Parsons Company Engineers/Constructors. Sulfur Recovery Study - Onshore Sour Gas Production Facilities, July 1981, and Small Sulfur Recovery Units - Onshore Sour Gas Production Facilities, April 1983. The Studies were conducted for TRW. The studies are presented in Appendix E and Appendix H, respectively of this document. Docket No. A-80-20-A, Entry II-A-16 and Entry II-A-20.

# 8. COST ANALYSIS

## 8.1 COST ANALYSIS OF REGULATORY ALTERNATIVES

This section summarizes the cost analysis data. It presents estimates of the fixed capital costs, annualized costs, capital charges, operating and maintenance costs as well as sulfur and steam credits for each of the six regulatory alternatives applied to all twenty-one model plants. Cost effectiveness and incremental cost effectiveness for each of the six regulatory alternatives applied to the model plants also are discussed. All of the estimates are based upon cost data developed in two studies conducted by a sulfur recovery equipment vendor.[1] Cost data pertinent to plants with 5 Mg/d and larger sulfur feed rates are presented in Appendix E. Appendix H summarizes the basic cost data for the plants with less than 5 Mg/d sulfur feed rates. The cost analysis and data presented in this chapter are applicable to sweetening and sulfur recovery technologies conducted onshore for the natural gas production industry. Cost estimates for sweetening and sulfur recovery technologies conducted offshore may vary widely from the data presented in this chapter. The regulatory alternatives, baseline controls (Regulatory Alternative I), and model plants are presented in Table 6-1.

### 8.1.1 New Facilities

In this section, the installed fixed-capital and annualized costs associated with each regulatory alternative (beyond the baseline controls) are presented for all twenty-one model plants. Sulfur dioxide ($SO_2$) emissions reductions and cost effectiveness for each regulatory alternative also are summarized for the model plants. All costs are for new facilities.

8.1.1.1 Capital Costs. The fixed capital costs that represent the initial investment for control equipment and installation for 51 different cases, 39 cases for >5 Mg/d sizes and 12 cases for $\leq$5 Mg/d sizes, which represent the size and acid gas ratio ranges in the industry, are presented in Appendix E and Appendix H of this document. The cases consider different model plant sizes (sulfur feed rates) and combinations of sulfur recovery/tail gas processes and acid gas $H_2S/CO_2$ ratios. The cost estimates for the model plant sizes greater than 5 Mg/d are based upon the following assumptions (Appendix E):

(1) Barometric pressure is 14.7 psia;

(2) Acid gases are available at 38°C (100°F) and 24.7 psia, saturated with water, and containing 0.5 mole percent methane (wet basis);

(3) Steam is produced at 250, 50, and 15 psig from the Claus stages;

(4) Thermal oxidizers are operated at 649°C (1,200°F) with 25 percent excess air (to reduce $H_2S$ content below 10 ppmv);

(5) For cases with waste heat boilers associated with thermal oxidizers, steam is produced at 250 psig for the 102 Mg/d (100 LT/D) cases and at both 600 and 250 psig for each of the 564 or 1,016 Mg/d (555 or 1,000 LT/D) cases;

(6) Treated, deaerated boiler feedwater is available at 320 psig and 110°C (230°F);

(7) Cooling water is available at 29°C (85°F) and is returned at 43°C (110°F);

(8) Incinerator stack heights vary from 30.5 to 183 m (100 to 600 ft) depending on the quantity of sulfur dioxide emissions; stack height was set to achieve approximately uniform levels of ground-level $SO_2$ concentrations;

(9) Investment costs are based on January 1981 Gulf Coast prices;

(10) In the BSR (hydrogenation) sections, steam is produced in the 102 Mg/d (100 LT/D) units at 50 psig and in the 564 or 1,016 Mg/d (555 or 1,000 LT/D) units at 450 and 50 psig;

(11) For the cases with 1,016 Mg/d (1,000 LT/D) sulfur input feeding acid gas with a 50/50 $H_2S/CO_2$ ratio, the plants are approximately the maximum economical size, and weaker gases would require building two trains. In order to keep all cases in only one train, the maximum sulfur input with 20/80 $H_2S/CO_2$ acid gas

feed is 564 Mg/d (555 LT/D). These plants are about the same physical size as the 1,016 Mg/d (1,000 LT/D) plants with 50/50 $H_2S/CO_2$ acid gas feed; and

(12) Initial charge of catalysts is included in the fixed capital costs.

The cost estimates for the model plant sizes smaller than 5 Mg/d are based upon the following assumptions (Appendix H):

(1) Barometric pressure is 14.7 psia;

(2) Acid gases are available at 120°F and 24.7 psia, saturated with water, and contain 0.5 volume % methane (wet basis) maximum.

(3) Investment costs in Table H-3 are based on February 1983 Gulf Coast prices and include the complete plants through the catalytic incinerator and 100-foot-high stack. (During the period from January 1981 to February 1983, Gulf Coast Petroleum Products Industry prices increased by about 9.6 percent (Based on Chemical Engineering Economic Index for the Petroleum Products Process Industry); therefore, the difference between the <5 Mg/d plant costs in 1981 dollars and the <5 Mg/d plant costs in 1983 dollars is considered to be insignificant.)

(4) The plants use a heating medium for preheaters and reheaters and a cooling medium by a cooler, but can also be recovered by circulating it through a reboiler on the regenerator in the adjacent amine unit. For these small plants, such a reboiler has not been included as the economics may be marginal.

The fixed capital costs of the regulatory alternatives for the model plants with sulfur feed rates above 5 Mg/d were taken from the data in Table E-28, Appendix E. These costs include the cost of the initial catalyst charge and a capital cost for stack sulfur dioxide monitoring equipment ($41,000). Those model plant/regulatory alternative combinations, for which case-specific costs were not estimated, were calculated through interpolation of the cost numbers based upon the model plant size for cases with identical sulfur recovery technology and acid gas $H_2S/CO_2$ ratio. The fixed-capital costs for the model plants with sulfur feed rates below 5 Mg/d were taken from the cost data presented in Table H-3, Appendix H, plus the cost for the initial catalyst charge (Table H-4) and a capital cost for stack sulfur dioxide monitoring equipment ($41,000). Those model plants (less than 5 Mg/d) for which case-specific costs were not presented, were calculated through interpolation/extrapolation of the cost numbers based upon the model plant size for cases with identical sulfur recovery technology and an acid gas $H_2S/CO_2$ ratio of 2/98 (i.e., the ratio that yields the highest cost for a given plant size in sulfur feed rate). The fixed-capital costs (incremental costs beyond the costs of the baseline control technology) for each new model plant/regulatory alternative combination are presented in Table 8-1. In addition, the fifth year (1987) aggregate fixed-capital costs for all plants projected for construction in the period 1983 through 1987 are listed at the base of each regulatory alternative column.

8.1.1.2 Annualized Costs. The annualized cost is the summation of capital charge cost, operating and maintenance costs, and other expenses minus credits, if any, from the use of by-product steam and the sale of elemental sulfur generated by the control processes. Credit components are deducted from the cost components. An operating schedule of 24 hours per day and 350 days per year was used to calculate cost and credit components.

Table 8-2 lists the 20 cost components, 7 credit components and the cost (and credit) factors used to estimate each component for the model plants with greater than 5 Mg/d sulfur rates. The cost (and credit) factors presented in Table 8-2 were used to calculate in a step-by-step fashion the annualized costs for the 39 cases for which case-specific fixed-capital costs were estimated.[4] These data (presented in Appendix E) were used as the basis for estimating annualized costs for specific new model plant/regulatory alternative combinations. The annualized costs for the 39 cases are presented in Table E-29, Appendix E. Those model plant/regulatory alternative combinations for which case-specific costs were not estimated, were calculated through interpolation of the cost numbers based upon the model plant size (sulfur feed rate) for cases with identical sulfur recovery technology and acid gas $H_2S/CO_2$ ratio.

## Table 8-1. FIXED-CAPITAL COSTS FOR EACH NEW MODEL PLANT/REGULATORY ALTERNATIVE COMBINATION

| Model plant # | Model plant size (sulfur feed rate) Mg/d | Projected number (five year total) | Regulatory alternative fixed-capital costs, 10³ $ (incremental costs beyond the baseline) | | | | | |
|---|---|---|---|---|---|---|---|---|
| | | | I (baseline control) | II | III | IV | V | VI |
| 1 | <0.1 (0.034 avg.) | 11 | 0 | 0 | 0 | 0 | 0 | 0 |
| 2 | 0.1 | 1 | 0 | 0 | 0 | 0 | 0 | 0 |
| 3 | 0.2 | 3 | 0 | 0 | 0 | 0 | 0 | 1,480 |
| 4 | 0.3 | 1 | 0 | 0 | 0 | 0 | 1,580 | 1,580 |
| 5 | 0.4 | 2 | 0 | 0 | 0 | 0 | 1,660 | 1,890 |
| 6 | 0.5 | 1 | 0 | 0 | 0 | 1,720 | 1,940 | 7,240 |
| 7 | 0.6 | 1 | 0 | 0 | 0 | 1,780 | 2,040 | 7,240 |
| 8 | 0.7 | 1 | 0 | 0 | 0 | 1,820 | 2,080 | 7,240 |
| 9 | 0.8 | 1 | 0 | 0 | 0 | 1,860 | 2,130 | 7,240 |
| 10 | 0.9 | 1 | 0 | 0 | 0 | 1,900 | 2,180 | 7,240 |
| 11 | 1.0 | 5 | 0 | 0 | 1,920 | 1,920 | 2,210 | 7,240 |
| 12 | 2.0 | 3 | 0 | 0 | 2,210 | 2,210 | 2,550 | 7,240 |
| 13 | 3.0 | 3 | 0 | 0 | 2,310 | 2,950 | 7,240 | 7,240 |
| 14 | 4.0 | 2 | 0 | 0 | 2,420 | 3,160 | 7,240 | 7,240 |
| 15 | 5.0 | 1 | 0 | 0 | 2,520 | 3,410 | 7,240 | 7,240 |
| 16 | 10 | 18 | 0 | 0 | 385 | 385 | 5,300 | 5,300 |
| 17 | 102 | 9 | 0 | 0 | 1,840 | 1,840 | 14,330 | 14,330 |
| 18 | 564 | 0 | 0 | 2,800 | 2,800 | 15,600 | 15,600 | 15,600 |
| 19 | 564 | 2 | 0 | 1,570 | 15,000 | 15,000 | 15,000 | 15,000 |
| 20 | 1,016 | 0 | 0 | 3,580 | 3,580 | 22,200 | 22,200 | 22,200 |
| 21 | 1,016 | 1 | 0 | 1,600 | 17,900 | 17,900 | 17,900 | 17,900 |
| | Fifth year total | 67 | 0 | 4,700 | 102,000 | 115,000 | 349,000 | 419,000 |

NOTE: All numbers rounded to three significant digits.

## Table 8-2. COMPONENTS OF ANNUALIZED COSTS AND FACTORS TO CALCULATE THESE COMPONENTS FOR THE MODEL PLANTS WITH GREATER THAN 5 Mg/d SULFUR FEED RATES

Annualized Costs = Operating Costs + General Expenses - Steam and Sulfur Credits.

A. Operating Costs:

1. Direct costs:

a. Utilities consumed: Based upon 24 h/d, 350 d/y schedule.

1. 600 PSIG steam, $15.98/Mg
2. 50 PSIG steam, $12.68/Mg
3. Treated boiler feed water, $3.31/Mg
4. Cooling water, \$13.21/$10^3 m^3$ circulated
5. Electric power, $0.05/kWh
6. Fuel gas, \$4.74/$10^9$J
7. Catalyst and chemical, ranging $2.27/day to $3,579/day depending upon size and technology.

b. Utilities generated (credits): Based upon 24 h/d, 350 d/y schedule.

8. 600 PSIG steam, $15.98/Mg
9. 450 PSIG steam, $15.43/Mg
10. 250 PSIG steam, $14.88/Mg
11. 50 PSIG steam, $12.68/Mg
12. 15 PSIG steam, $9.92/Mg
13. Steam condensate, $2.76/Mg
14. Sulfur recovered, $98.42/Mg

c. Operating labor: Based on 8 hr/operator (or supervisor), 3 shifts/d, 365 d/y schedule.

15. Operators per shift: ($14.50/h rate)

| Cases (For cases, refer to Table E-29) | Number per shift |
|---|---|
| 1 and 2 | negligible |
| 3,4,5,9,10,14,15,19,20 | 0.75 |
| 6,7,8,11,12,13,16,17,18,29,30,31 | 1.25 |
| All others | 2.25 |

16. Supervisors per shift: ($18.80/h rate)
All cases except 1 and 2 0.25

d. Maintenance and repair:

17. Labor, 1.5% of fixed-capital costs

18. Materials, 2% of fixed-capital costs

(continued)

Table 8-2. Concluded

---

e. Operating supplies:

19. Operating supplies, 10% of operating labor

f. Laboratory charges

20. Laboratory charges, 10% of operating labor

g. Continuous stack $SO_2$ mass emission monitor system

21. Operating costs, $37,900/y

2. Fixed charges:

22. Capital charges = fixed-capital costs x $\frac{i\,(1+i)^n}{(1+i)^n-1}$

= 0.11746 x fixed-capital costs

where, $i$ = 10% (interest rate)
$n$ = 20 years (equipment life)

23. Local taxes, 1% of fixed-capital costs

24. Insurance, 0.6% of fixed-capital costs

3. Plant overhead costs:

25. Plant overhead, 25% of operating labor plus maintenance and repair

B. General Expenses:

1. Administrative costs:

26. Administration, 1% of total operating costs

2. Distribution and selling costs:

27. Distribution and selling, 1% of total operating costs

---

Source: (i) The factors to calculate components 20, 23, 26, and 27 were taken from Reference 2.

(ii) The factor to calculate component 22 was taken from Reference 3.

(iii) The factor to calculate component 21 was based upon information from Western Research and Development, Calgary, Canada.

(iv) The factors to calculate the remaining components were taken from Reference 1 (Appendix E).

Table 8-3 lists the 19 cost components and the factors used to estimate each component for the model plants with less than 5 Mg/d sulfur feed rates. The cost factors presented in Table 8-3 were used to calculate in a step-by-step fashion the annualized costs for the 12 cases for which case-specific fixed-capital costs were estimated.[5] These data (presented in Appendix H) were used as the basis for estimating annualized costs for specific new model plant/regulatory alternative combinations. The annualized costs for the 12 cases are presented in Table H-7, Appendix H. Those model plant/regulatory alternative combinations for which case-specific costs were not estimated, were calculated through interpolation/extrapolation of the cost numbers based upon the model plant size for cases with identical sulfur recovery technology and an acid gas $H_2S/CO_2$ ratio of 2/98 (i.e., the ratio that yields the highest cost for a given plant size in sulfur feed rate).

The annualized costs (incremental costs beyond the baseline) for each model plant/regulatory alternative combination are presented in Table 8-4. In addition, the fifth year (1987) aggregate annualized costs for all plants projected for construction in the period 1983 through 1987 are listed at the base of each regulatory alternative column. Incremental costs for capital charge cost, operating and maintenance costs and other expenses and incremental credits from steam and sulfur generated for each model plant/regulatory combination are presented in Table 8-5, 8-6 and 8-7, respectively. Sulfur credits were assumed to be zero for the model plants with sulfur feed rates below 5 Mg/d. For this plant size range it also was assumed that a sulfur disposal cost of $25/LT was incurred.

8.1.1.3 Cost Effectiveness. To determine cost effectiveness for each individual model plant in Regulatory Alternatives II through VI, the annualized costs (from Table 8-4) associated with each of these regulatory alternatives were subtracted from the annualized costs associated with the baseline controls (Regulatory Alternative I) and then divided by the $SO_2$ emissions reductions (calculated in Chapter 7) for each of the regulatory alternatives (II through VI) beyond the Regulatory Alternative I emissions. Cost effectiveness and incremental

Table 8-3. COMPONENTS OF ANNUALIZED COSTS AND FACTORS TO CALCULATE THESE COMPONENTS FOR THE MODEL PLANTS WITH LESS THAN 5 Mg/d SULFUR FEED RATES

---

Annualized Costs = Operating Costs + General Expenses

A. Operating Costs:

1. Direct costs:

a. Utilities consumed: Based on 24 h/d, 350 d/y schedule

1. Electric power, \$0.05/kWh
2. Fuel gas, \$4.74/$10^9$J
3. Catalysts, ranging \$1.48/day to \$56.4/day depending upon size and technology

b. Utilities generated (credits): Based upon 24 h/d, 350 d/y schedule

4. Assumed zero credits for the model plants with less than 5 Mg/d sulfur feed rates

c. Operating labor: Based on 8 hr/operator (or supervisor), 3 shifts/d, 365 d/y schedule

5. Operators per shift: (\$14.50/h rate)

| Cases (For cases, refer to Table H-5) | Number per shift |
|---|---|
| 1 through 12 (all the cases) | 0.75 |

6. Supervisors per shift: (\$18.80/h rate)

| | |
|---|---|
| All the cases | 0.25 |

d. Maintenance and repair:

7. Labor, 1.5% of fixed capital costs

8. Materials, 2% of fixed-capital costs

e. Operating supplies:

9. Operating supplies, 10% of operating labor

f. Laboratory charges:

10. Laboratory charges, 10% of operating labor

g. Continuous stack $SO_2$ mass emission monitor system:

11. Operating costs, \$37,900/y

h. Storage and disposal of recovered sulfur:

12. Storage, disposal costs, \$0.25/Mg/mile (100 miles average distance for disposal assumed)

---

(continued)

Table 8-3. Concluded

2. Fixed charges:

   13.

   $$\text{Capital charges} = \text{fixed-capital costs} \times \frac{i\,(1 + i)^n}{(1 + i)^n - 1}$$

   $$= 0.11746 \times \text{fixed-capital costs}$$

   where, $i$ = 10% (interest rate)
   $n$ = 20 years (equipment life)

   14. Local taxes, 1% of fixed-capital costs

   15. Insurance, 0.6% of fixed-capital costs

3. Plant overhead costs:

   16. Plant overhead, 25% of operating labor plus maintenance and repair

B. General Expenses:

1. Administrative costs:

   17. Administration, 1% of total operating costs

2. Distribution and selling costs:

   18. Distribution and selling, 1% of total operating costs

3. Contingency costs:

   19. Contingency factor, 5% of total operating costs

---

Source:

(i) The factors to calculate components 10, 14, 17, 18, and 19 were taken from Reference 2.

(ii) The factor to calculate component 13 was taken from Reference 3.

(iii) The factor to calculate component 11 was based upon information from Western Research and Development, Calgary, Canada.

(iv) The factors to calculate the remaining components were taken from Reference 1 (Appendix H).

## Table 8-4. ANNUALIZED COSTS FOR EACH NEW MODEL PLANT/REGULATORY ALTERNATIVE COMBINATION

| Model plant # | Model plant size (sulfur feed rate) Mg/d | Projected number (five year total) | Regulatory alternative annualized costs, 10³ $/yr (incremental costs beyond the baseline) | | | | | |
|---|---|---|---|---|---|---|---|---|
| | | | I (baseline control) | II | III | IV | V | VI |
| 1 | <0.1[a] | 11 | 0 | 0 | 0 | 0 | 0 | 0 |
| 2 | 0.1[a] | 1 | 0 | 0 | 0 | 0 | 0 | 0 |
| 3 | 0.2[a] | 3 | 0 | 0 | 0 | 0 | 0 | 434 |
| 4 | 0.3[a] | 1 | 0 | 0 | 0 | 0 | 446 | 446 |
| 5 | 0.4[a] | 2 | 0 | 0 | 0 | 0 | 461 | 640 |
| 6 | 0.5[a] | 1 | 0 | 0 | 0 | 475 | 648 | 1,100 |
| 7 | 0.6[a] | 1 | 0 | 0 | 0 | 488 | 656 | 1,100 |
| 8 | 0.7[a] | 1 | 0 | 0 | 0 | 501 | 664 | 1,100 |
| 9 | 0.8[a] | 1 | 0 | 0 | 0 | 515 | 670 | 1,100 |
| 10 | 0.9[a] | 1 | 0 | 0 | 0 | 528 | 678 | 1,100 |
| 11 | 1.0[a] | 5 | 0 | 0 | 541 | 541 | 685 | 1,100 |
| 12 | 2.0[a] | 3 | 0 | 0 | 674 | 674 | 759 | 1,100 |
| 13 | 3.0[a] | 3 | 0 | 0 | 765 | 833 | 1,100 | 1,100 |
| 14 | 4.0[a] | 2 | 0 | 0 | 856 | 906 | 1,100 | 1,100 |
| 15 | 5.0 | 1 | 0 | 0 | 448 | 483 | 855 | 855 |
| 16 | 10 | 18 | 0 | 0 | 103 | 103 | 722 | 722 |
| 17 | 102 | 9 | 0 | 0 | 593 | 593 | 8,200 | 8,200 |
| 18 | 564 | 0 | 0 | 338 | 338 | 10,500 | 10,500 | 10,500 |
| 19 | 564 | 2 | 0 | 192 | 4,260 | 4,260 | 4,260 | 4,260 |
| 20 | 1,016 | 0 | 0 | 429 | 429 | 16,600 | 16,600 | 16,600 |
| 21 | 1,016 | 1 | 0 | 200 | 5,770 | 5,770 | 5,770 | 5,770 |
| | Fifth year total | 67 | 0 | 584 | 30,700 | 33,500 | 118,000 | 125,000 |

[a] Sulfur credits were assumed to be zero for the model plants with sulfur feed rates below 5 Mg/d.

NOTE: All numbers rounded to three significant digits.

Table 8-5. CAPITAL CHARGE COSTS FOR EACH NEW MODEL PLANT/REGULATORY ALTERNATIVE COMBINATION

| Model plant # | Model plant size (sulfur feed rate) Mg/d | Projected number (five year total) | Regulatory alternative capital charge costs, $10^3$ \$/yr (incremental costs beyond the baseline) | | | | | |
|---|---|---|---|---|---|---|---|---|
| | | | I (baseline control) | II | III | IV | V | VI |
| 1 | <0.1 (0.034 avg.) | 11 | 0 | 0 | 0 | 0 | 0 | 0 |
| 2 | 0.1 | 1 | 0 | 0 | 0 | 0 | 0 | 0 |
| 3 | 0.2 | 3 | 0 | 0 | 0 | 0 | 0 | 174 |
| 4 | 0.3 | 1 | 0 | 0 | 0 | 0 | 186 | 186 |
| 5 | 0.4 | 2 | 0 | 0 | 0 | 0 | 195 | 222 |
| 6 | 0.5 | 1 | 0 | 0 | 0 | 202 | 228 | 850 |
| 7 | 0.6 | 1 | 0 | 0 | 0 | 209 | 240 | 850 |
| 8 | 0.7 | 1 | 0 | 0 | 0 | 214 | 244 | 850 |
| 9 | 0.8 | 1 | 0 | 0 | 0 | 218 | 250 | 850 |
| 10 | 0.9 | 1 | 0 | 0 | 0 | 223 | 256 | 850 |
| 11 | 1.0 | 5 | 0 | 0 | 226 | 226 | 260 | 850 |
| 12 | 2.0 | 3 | 0 | 0 | 260 | 260 | 264 | 850 |
| 13 | 3.0 | 3 | 0 | 0 | 271 | 347 | 850 | 850 |
| 14 | 4.0 | 2 | 0 | 0 | 284 | 371 | 850 | 850 |
| 15 | 5.0 | 1 | 0 | 0 | 296 | 401 | 850 | 850 |
| 16 | 10 | 18 | 0 | 0 | 45 | 45 | 618 | 618 |
| 17 | 102 | 9 | 0 | 0 | 216 | 216 | 1,680 | 1,680 |
| 18 | 564 | 0 | 0 | 331 | 331 | 1,830 | 1,830 | 1,830 |
| 19 | 564 | 2 | 0 | 184 | 1,760 | 1,760 | 1,760 | 1,760 |
| 20 | 1,016 | 0 | 0 | 420 | 420 | 2,600 | 2,600 | 2,600 |
| 21 | 1,016 | 1 | 0 | 191 | 2,100 | 2,110 | 2,110 | 2,110 |
| | Fifth year total | 67 | 0 | 559 | 12,000 | 13,500 | 40,900 | 49,200 |

NOTE: All numbers rounded to three significant digits.

## Table 8-6. OPERATING AND MAINTENANCE COSTS AND OTHER EXPENSES FOR EACH NEW MODEL PLANT/REGULATORY ALTERNATIVE COMBINATION

| Model plant # | Model plant size (sulfur feed rate) Mg/d | Projected number (five year total) | Regulatory alternative operating and maintenance costs and other expenses, $10^3$ \$/yr (incremental costs beyond the baseline) | | | | | |
|---|---|---|---|---|---|---|---|---|
| | | | I (baseline control) | II | III | IV | V | VI |
| 1 | <0.1 (0.034 avg.) | 11 | 0 | 0 | 0 | 0 | 0 | 0 |
| 2 | 0.1 | 1 | 0 | 0 | 0 | 0 | 0 | 0 |
| 3 | 0.2 | 3 | 0 | 0 | 0 | 0 | 0 | 260 |
| 4 | 0.3 | 1 | 0 | 0 | 0 | 0 | 260 | 260 |
| 5 | 0.4 | 2 | 0 | 0 | 0 | 0 | 266 | 418 |
| 6 | 0.5 | 1 | 0 | 0 | 0 | 273 | 420 | 250 |
| 7 | 0.6 | 1 | 0 | 0 | 0 | 279 | 416 | 250 |
| 8 | 0.7 | 1 | 0 | 0 | 0 | 287 | 420 | 250 |
| 9 | 0.8 | 1 | 0 | 0 | 0 | 297 | 420 | 250 |
| 10 | 0.9 | 1 | 0 | 0 | 0 | 305 | 422 | 250 |
| 11 | 1.0 | 5 | 0 | 0 | 315 | 315 | 425 | 250 |
| 12 | 2.0 | 3 | 0 | 0 | 414 | 414 | 495 | 250 |
| 13 | 3.0 | 3 | 0 | 0 | 494 | 486 | 250 | 250 |
| 14 | 4.0 | 2 | 0 | 0 | 572 | 535 | 250 | 250 |
| 15 | 5.0 | 1 | 0 | 0 | 292 | 225 | 254 | 254 |
| 16 | 10 | 18 | 0 | 0 | 71 | 71 | 467 | 467 |
| 17 | 102 | 9 | 0 | 0 | 493 | 493 | 7,720 | 7,720 |
| 18 | 564 | 0 | 0 | 767 | 767 | 9,228 | 9,228 | 9,228 |
| 19 | 564 | 2 | 0 | 649 | 2,959 | 2,959 | 2,959 | 2,959 |
| 20 | 1,016 | 0 | 0 | 1,169 | 1,169 | 15,140 | 15,140 | 15,140 |
| 21 | 1,016 | 1 | 0 | 963 | 4,629 | 4,629 | 4,629 | 4,629 |
| Fifth year total | | 67 | 0 | 2,260 | 22,000 | 23,300 | 96,400 | 95,100 |

NOTE: All numbers rounded to three significant digits.

Table 8-7. SULFUR AND STEAM CREDITS FOR EACH NEW MODEL PLANT/REGULATORY ALTERNATIVE COMBINATION

| Model plant[a] # | Model plant size (sulfur feed rate) Mg/d | Projected number (five year total) | Regulatory alternative steam and sulfur credits, $10^3$ \$/yr (incremental credits beyond the baseline) | | | | | |
|---|---|---|---|---|---|---|---|---|
| | | | I (baseline control) | II | III | IV | V | VI |
| 1 | <0.1 (0.034 avg.) | 11 | (0) | (0) | (0) | (0) | (0) | (0) |
| 2 | 0.1 | 1 | (0) | (0) | (0) | (0) | (0) | (0) |
| 3 | 0.2 | 3 | (0) | (0) | (0) | (0) | (0) | (0) |
| 4 | 0.3 | 1 | (0) | (0) | (0) | (0) | (0) | (0) |
| 5 | 0.4 | 2 | (0) | (0) | (0) | (0) | (0) | (0) |
| 6 | 0.5 | 1 | (0) | (0) | (0) | (0) | (0) | (0) |
| 7 | 0.6 | 1 | (0) | (0) | (0) | (0) | (0) | (0) |
| 8 | 0.7 | 1 | (0) | (0) | (0) | (0) | (0) | (0) |
| 9 | 0.8 | 1 | (0) | (0) | (0) | (0) | (0) | (0) |
| 10 | 0.9 | 1 | (0) | (0) | (0) | (0) | (0) | (0) |
| 11 | 1.0 | 5 | (0) | (0) | (0) | (0) | (0) | (0) |
| 12 | 2.0 | 3 | (0) | (0) | (0) | (0) | (0) | (0) |
| 13 | 3.0 | 3 | (0) | (0) | (0) | (0) | (0) | (0) |
| 14 | 4.0 | 2 | (0) | (0) | (0) | (0) | (0) | (0) |
| 15 | 5.0 | 1 | (0) | (0) | (140) | (143) | (249) | (249) |
| 16 | 10 | 18 | (0) | (0) | (13) | (13) | (363) | (363) |
| 17 | 102 | 9 | (0) | (0) | (116) | (116) | (1,200) | (1,200) |
| 18 | 564 | 0 | (0) | (760) | (760) | (558) | (558) | (558) |
| 19 | 564 | 2 | (0) | (641) | (459) | (459) | (459) | (459) |
| 20 | 1,016 | 0 | (0) | (1,160) | (1,160) | (1,140) | (1,140) | (1,140) |
| 21 | 1,016 | 1 | (0) | (954) | (959) | (959) | (959) | (959) |
| | Fifth year total | 67 | (0) | (2,240) | (3,300) | (3,290) | (19,500) | (19,500) |

[a] Sulfur credits assumed to be zero for the model plants with sulfur feed rates below 5 Mg/d.

NOTE: All numbers rounded to three significant digits.

cost effectiveness were calculated for each model plant size in each regulatory alternative. The cost effectiveness is the annualized cost (beyond the baseline control) per megagram of $SO_2$ emissions reduction achieved (beyond the baseline control). The incremental cost effectiveness of Alternative III is the additional annualized cost of Alternative III over Alternative II divided by the additional $SO_2$ emissions reduction achieved by Alternative III over Alternative II. This same approach is used to calculate the incremental cost effectiveness for Alternative IV, V and VI. The cost effectiveness and incremental cost effectiveness values for each regulatory alternative/model plant combination is summarized in Table 8-8.

For the model plants with sulfur feed rates below 5 Mg/d, cost-effectiveness values were calculated for the Recycle Selectox 2-stage and 3-stage processes. Table 8-9 presents cost effectiveness values for the Recycle Selectox 2-stage process and Table 8-10 for the Recycle Selectox 3-stage process. Table 8-11 presents incremental cost effectiveness values between the Recycle Selectox 2-stage and 3-stage processes for these small size model plants.

No additional costs are incurred because affected facilities are expected already to be in compliance with any existing applicable regulations. No other regulatory requirements are being imposed on the affected facilities.

### 8.1.2 Modified or Reconstructed Facilities

As stated in Chapter 5 of this document, no physical or operational changes in the industry are anticipated that would qualify an existing facility as "modification." Also, the cost attributable to any replacement would not exceed 50 percent of the cost to construct a comparable, entirely new facility. Consequently, no situations are anticipated that would constitute a "reconstruction" to an exisiting facility. Therefore, the costs for "modification" or "reconstruction" have not been developed.

## Table 8-8. COST EFFECTIVENESS[a] AND INCREMENTAL COST EFFECTIVENESS[b] FOR EACH MODEL PLANT/REGULATORY ALTERNATIVE COMBINATION (\$/Mg $SO_2$)

| Model plant | Sulfur feed rate[c], Mg/D | I (baseline control) | ΔCE | II | ΔCE | III | ΔCE | IV | ΔCE | V | ΔCE | VI |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| 1 | <0.1[c] (0.034 avg.) | — | — | — | — | — | | — | | — | | — |
| 2 | 0.1[c] | — | — | — | — | — | | — | | — | | — |
| 3 | 0.2[c] | — | — | — | — | — | | — | | — | | 4,130 |
| 4 | 0.3c | — | — | — | — | — | | — | | 2,840 | — | 2,840 |
| 5 | 0.4[c] | — | — | — | — | — | | — | | 2,200 | 30,600 | 2,980 |
| 6 | 0.5[c] | — | — | — | — | — | | 1,810 | 23,700 | 2,410 | 44,800 | 3,940 |
| 7 | 0.6[c] | — | — | — | — | — | | 1,550 | 19,100 | 2,030 | 36,700 | 3,280 |
| 8 | 0.7[c] | — | — | — | — | — | | 1,370 | 15,900 | 1,760 | 30,900 | 2,810 |
| 9 | 0.8[c] | — | — | — | — | — | | 1,230 | 13,200 | 1,550 | 26,700 | 2,460 |
| 10 | 0.9[c] | — | — | — | — | — | | 1,120 | 11,400 | 1,400 | 23,300 | 2,190 |
| 11 | 1.0[c] | — | — | — | 1,030 | 1,030 | — | 1,030 | 9,840 | 1,270 | 20,600 | 1,970 |
| 12 | 2.0[c] | — | — | — | 642 | 642 | — | 642 | 2,910 | 703 | 8,460 | 985 |
| 13 | 3.0[c] | — | — | — | 487 | 487 | 1,550 | 514 | 4,420 | 656 | — | 656 |
| 14 | 4.0[c] | — | — | — | 408 | 408 | 854 | 421 | 2,410 | 492 | — | 492 |
| 15 | 5.0[d] | — | — | — | 171 | 171 | 478 | 180 | 3,690 | 306 | — | 306 |
| 16 | 10[d] | — | — | — | 630 | 630 | — | 630 | 6,250 | 2,750 | — | 2,750 |
| 17 | 102[d] | — | — | — | 350 | 350 | — | 350 | 7,240 | 2,990 | — | 2,990 |
| 18 | 564 (50/50) | — | 36 | 36 | — | 36 | 1,680 | 680 | — | 680 | — | 680 |
| 19 | 564 (80/20) | — | 26 | 26 | 690 | 320 | — | 320 | — | 320 | — | 320 |
| 20 | 1,016 (50/50) | — | 25 | 25 | — | 25 | 1,500 | 600 | — | 600 | — | 600 |
| 21 | 1,016 (80/20) | — | 15 | 15 | 525 | 240 | — | 240 | — | 240 | — | 240 |

[a] Cost effectiveness = Annualized costs of the given regulatory alternative minus annualized costs of the baseline control divided by the annual long tons $SO_2$ emissions further reduced from the baseline control.

[b] Incremental cost effectiveness (ΔCE) = Difference between the annualized costs of the given and the previous regulatory alternative divided by the difference in the annual long tons $SO_2$ emissions reduction between the same alternatives.

[c] Covers ratio range of less than 2/98 to over 12.5/87.5.

[d] Covers the entire ratio range from less than 12.5/87.5 to over 80/20.

Table 8-9. COST EFFECTIVENESS OF RECYCLE SELECTOX 2-STAGE PROCESS

| Model plant | Sulfur feed rate, LT/D | Annualized costs of incinerator, \$/year (baseline controls) | Incremental annualized costs (beyond the baseline controls) of Recycle Selectox 2-stage \$/year[a] | Reduction in $SO_2$ emissions from the baseline control Mg $SO_2$/year[b] | Cost effectiveness, \$/Mg $SO_2$ |
|---|---|---|---|---|---|
| 1 | <0.1<br>(0.034 avg.) | 154,000 | 408,000 | 17.8 | 22,900 |
| 2 | 0.1 | 154,000 | 421,000 | 52.4 | 8,030 |
| 3 | 0.2 | 154,000 | 434,000 | 105 | 4,130 |
| 4 | 0.3 | 154,000 | 446,000 | 157 | 2,840 |
| 5 | 0.4 | 154,000 | 461,000 | 210 | 2,200 |
| 6 | 0.5 | 154,000 | 475,000 | 262 | 1,810 |
| 7 | 0.6 | 154,000 | 488,000 | 314 | 1,550 |
| 8 | 0.7 | 154,000 | 501,000 | 367 | 1,370 |
| 9 | 0.8 | 154,000 | 515,000 | 419 | 1,230 |
| 10 | 0.9 | 154,000 | 528,000 | 471 | 1,120 |
| 11 | 1.0 | 154,000 | 541,000 | 525 | 1,030 |
| 12 | 2.0 | 154,000 | 674,000 | 1,050 | 642 |
| 13 | 3.0 | 154,000 | 765,000 | 1,570 | 487 |
| 14 | 4.0 | 154,000 | 856,000 | 2,100 | 408 |
| 15 | 5.0 | 154,000 | 448,000 | 2,620 | 171 |

[a] Monitoring and recordkeeping costs are included. Sulfur credits are not accounted for in the model plants with sulfur feed rates below 5 LT/D. Instead, storage and disposal costs of recovered liquid sulfur are accounted for. Sulfur (and steam) credits are accounted for in the model plants with sulfur feed rates 5 LT/D or larger. Sulfur credit from sale at \$100/LT.

[b] Based upon end-of-run sulfur recovery efficiency of 74.83 percent of Recycle Selectox 2-stage at 2 mole percent $H_2S$ in the acid gas stream.

Table 8-10. COST EFFECTIVENESS OF RECYCLE SELECTOX 3-STAGE PROCESS

| Model plant | Sulfur feed rate, LT/D | Annualized costs of incinerator, \$/year (baseline controls) | Incremental annualized costs (beyond the baseline controls) of Recycle Selectox 3-stage \$/year[a] | Reduction in $SO_2$ emissions from the baseline control Mg $SO_2$/year[b] | Cost effectiveness, \$/Mg $SO_2$ |
|---|---|---|---|---|---|
| 1 | <0.1<br>(0.034 avg.) | 154,000 | 611,000 | 18.3 | 33,400 |
| 2 | 0.1 | 154,000 | 618,000 | 53.8 | 11,500 |
| 3 | 0.2 | 154,000 | 626,000 | 108 | 5,800 |
| 4 | 0.3 | 154,000 | 633,000 | 162 | 3,910 |
| 5 | 0.4 | 154,000 | 640,000 | 215 | 2,980 |
| 6 | 0.5 | 154,000 | 648,000 | 269 | 2,410 |
| 7 | 0.6 | 154,000 | 656,000 | 323 | 2,030 |
| 8 | 0.7 | 154,000 | 664,000 | 377 | 1,760 |
| 9 | 0.8 | 154,000 | 670,000 | 431 | 1,550 |
| 10 | 0.9 | 154,000 | 678,000 | 485 | 1,400 |
| 11 | 1.0 | 154,000 | 685,000 | 538 | 1,270 |
| 12 | 2.0 | 154,000 | 759,000 | 1,080 | 703 |
| 13 | 3.0 | 154,000 | 833,000 | 1,620 | 514 |
| 14 | 4.0 | 154,000 | 906,000 | 2,150 | 421 |
| 15 | 5.0 | 154,000 | 483,000 | 2,690 | 180 |

[a] Monitoring and recordkeeping costs are included. Sulfur credits are not accounted for in the model plants with sulfur feed rates below 5 LT/D. Instead, storage and disposal costs of recovered liquid sulfur are accounted for. Sulfur (and steam) credits are accounted for in the model plants with sulfur feed rates 5 LT/D or larger. Sulfur credit from sale at \$100/LT.

[b] Based upon end-of-run sulfur recovery efficiency of 76.92 percent of Recycle Selectox 3-stage at 2 mole percent $H_2S$ in the acid gas stream.

Table 8-11. COST EFFECTIVENESS OF RECYCLE SELECTOX 2-STAGE AND RECYCLE SELECTOX 3-STAGE PROCESSES AND THEIR INCREMENTAL COST EFFECTIVENESS

| Model plant | Sulfur feed rate, LT/D | Cost effectiveness[a] of Recycle Selectox 2-stage, \$/LT $SO_2$ | Cost effectiveness[a] of Recycle Selectox 3-stage, \$/LT $SO_2$ | Incremental cost effectivness[b] from Recycle Selectox 2-stage to 3-stage, \$/LT $SO_2$ |
|---|---|---|---|---|
| 1 | <0.1<br>(0.034 avg.) | 22,900 | 33,400 | 408,000 |
| 2 | 0.1 | 8,030 | 11,500 | 135,000 |
| 3 | 0.2 | 4,130 | 5,800 | 65,600 |
| 4 | 0.3 | 2,840 | 3,910 | 42,600 |
| 5 | 0.4 | 2,200 | 2,980 | 30,600 |
| 6 | 0.5 | 1,810 | 2,410 | 23,700 |
| 7 | 0.6 | 1,550 | 2,030 | 19,100 |
| 8 | 0.7 | 1,370 | 1,760 | 15,900 |
| 9 | 0.8 | 1,230 | 1,550 | 13,200 |
| 10 | 0.9 | 1,120 | 1,400 | 11,400 |
| 11 | 1.0 | 1,030 | 1,270 | 9,840 |
| 12 | 2.0 | 642 | 703 | 2,910 |
| 13 | 3.0 | 487 | 514 | 1,550 |
| 14 | 4.0 | 408 | 421 | 854 |
| 15 | 5.0 | 171 | 180 | 478 |

[a] Cost effectiveness = Annualized costs of the given regulatory alternative minus annualized costs of the baseline control divided by the annual long tons $SO_2$ emissions further reduced from the baseline control.

[b] Incremental cost effectiveness ($\Delta$CE) = Difference between the annualized costs of the given and the previous regulatory alternative divided by the difference in the annual long tons $SO_2$ emissions reduction between the same alternatives.

## 8.2 REFERENCES FOR CHAPTER 8

1. The Ralph M. Parsons Company, Engineers/Constructors. Sulfur Recovery Studies—Onshore Sour Natural Gas Production Facilities. The studies were conducted for TRW. July 1981 and April 1983. The studies are presented in Appendices E and H. Refer to Appendix E and Appendix H of this document. Docket No. A-80-20-A, Entry II-A-16 and Entry II-A-20.

2. Peters, M.S., and K.D. Timmerhaus. Plant Design and Economics for Chemical Engineers, Second Edition. McGraw-Hill Book Company, New York, 1968. Docket No. A-80-20-A, Entry II-I-74.

3. U.S. EPA, Economics Analysis Branch, Capital and Operating Costs of Selected Air Pollution Control Systems, EPA 450/5-80-002, EPA Contract No. 68-02-2899, R. B. Neveril, GARD, Incorporated, December 1978. Docket No. A-80-20-A, Entry II-A-11.

4. Docket No. A-80-20-A, Entry II-B-29, Methodology for converting Parsons' costs data (in Appendix E) to the annualized costs presented in Chapter 8 of the Background Information Document. Step-by-step sample calculations to calculate annualized costs also are included.

5. Docket No. A-80-20-A, Entry II-B-44, Methodology for converting Parsons' costs data (in Appendix H) to the annualized costs presented in Chapter 8 of the Background Information Document. Step-by-step sample calculations to calculate annualized costs also are included.

# 9. ECONOMIC ANALYSIS OF THE REGULATORY ALTERNATIVES

## 9.1 INDUSTRY PROFILE

This section describes the general business and economic conditions of the natural gas production industry. The primary focus of the discussion is on the production facilities that recover elemental liquid sulfur from acid gas.

Growth projections for the year 1987, five years after a proposal date of 1982 for the NSPS, were developed to illustrate the future trend of the industry. The profile and the projections are presented to aid in the determination of economic, energy, and environmental impacts of the proposed standard.

### 9.1.1 The Natural Gas Production Industry

The natural gas system in the United States consists of producers, processors, dealers, interstate and intrastate pipelines, distributors and consumers.

The production industry includes hundreds of firms engaged in the exploration, drilling, producing and processing of natural gas although a relatively small number of companies dominate the industry. The American Association of Petroleum Geologists (AAPG) states that the 16 largest firms in the industry found 40.3 percent of 41.3 trillion cubic feet of natural gas discovered during the period from 1969 to 1978. The AAPG also states that the 16 largest companies accounted for approximately 60 percent of industry expenditures for geological and geophysical information and lease acquisition. However, these large companies spend almost twice as much money as smaller firms on predrilling exploration and one-half as much as the others on wildcat drilling.[1]

Some companies engage in the operations of drilling, producing and processing of natural gas. Additionally, if the natural gas is sour, companies incorporate a sweetening operation that removes $H_2S$ and $CO_2$ from the sour natural gas. Sulfur can then be removed from the separated acid gas. Other companies drill for and produce sour natural gas, but contract with companies owning sweetening facilities to perform the sweetening and sulfur recovery operations for them. Individual ownership or joint ownership in one or all of these operations exists in this industry. In the 1960-1979 period, about 19 percent of onshore natural gas was estimated to have undergone sweetening. Roughly one third of this, or 6.3 percent, has been associated with natural gas sulfur recovery. The API Gas Plant Survey data (presented in Appendix G) indicated that in 1982, about 24 percent of the onshore natural gas was sour and 32 percent of this, or 7.7 percent of the onshore natural gas, was associated with sulfur recovery. Historically the natural gas that has been produced offshore has been sweet. For the growth projections EPA assumed that during the projecting period the natural gas produced offshore would be sweet.

About two-thirds of all natural gas is transmitted in pipelines across state lines to be sold in various metropolitan areas. The remainder is sold in intrastate markets. Approximately 100 pipeline companies operate the interstate pipeline network. This sector of the industry, more so than the production sector, tends to be dominated by large companies. In 1971, the four largest pipeline companies accounted for 35 percent of the total interstate pipeline volume, while the 20 largest companies transported over 93 percent of the gas.

Companies involved in final distribution of the gas constitute the least concentrated sector of the industry. Over 1,600 companies buy gas from pipelines and distribute it to various communities. Because they operate in different service areas, these companies rarely compete with one another, and are often regulated by state or local agencies.

There is some vertical integration in the industry with pipeline companies often owning producing wells. However, few companies engage in production, transmission and distribution of the gas. In contrast, horizontal integration is quite extensive. In the production sector, many companies produce crude oil and natural gas liquids in addition to natural gas although no one company predominates. In addition, many have investments in coal, oil shale, synfuels and mineral industries.

9.1.1.1 Number, Size and Location of Natural Gas Wells. The American Gas Association (AGA) reports that in 1979 over 20 trillion cubic feet of natural gas was produced by nearly 170,000 producing gas wells located in 30 states. The leading producing states are Texas, Louisiana, Oklahoma, New Mexico and Kansas. The number of wells and production by state are shown in Table 9-1 which lists the states in rank order according to average gas well size (flow rate) based on revised 1979 figures from AGA.[2] The largest wells on average are located in Alaska and produce over 10,000 Mcfd. The smallest wells on average are located in Oregon, Indiana and Maryland where the average gas wells produce less than 10 Mcfd. Nonassociated (non-oil producing) gas wells, which produce less than a maximum of 60 Mcfd, are defined by the Natural Gas Policy Act to be stripper gas wells.

Based on the figures in Table 9-1, the distribution of domestic gas wells and production by well size is estimated to be as follows.

| Well Size (Mcfd) | Percent of Total Wells | Percent of Total Production |
|---|---|---|
| Less than 100 | 43 | 2 |
| 100-400 | 25 | 20 |
| 400-800 | 22 | 36 |
| More than 800 | 10 | 42 |
| | 100 | 100 |

At the low end of the well size distribution, about 43 percent of all producing gas wells supply two percent of total production. At the high end, about ten percent of the wells supply 42 percent of total net wellhead production from dry gas and condensate wells. Excluded from this distribution are oil wells which also produce gas.

9.1.1.2 Natural Gas Production Costs from New Wells. The cost of producing natural gas is highly variable and depends on many factors such as drilling depth, probability of drilling a producing well, well flow rates, etc. Among these factors, drilling costs are the most important.

Table 9-1. NUMBER, AVERAGE SIZE, AND TOTAL PRODUCTION OF PRODUCING GAS WELLS, BY STATE (REVISED 1979 FIGURES)

| State | Average gas well size Mcfd[a] | Number of wells | Cumulative wells | Total state production (MMcfy) | Cumulative production (MMcfy) |
|---|---|---|---|---|---|
| Oregon | 1 | 5 | 5 | 2 | 2 |
| Indiana | 1 | 655 | 660 | 350 | 352 |
| Maryland | 6 | 12 | 672 | 28 | 380 |
| Pennsylvania | 13 | 19,800 | 20,472 | 96,313 | 96,693 |
| Tennessee | 16 | 184 | 20,656 | 1,067 | 97,760 |
| West Virginia | 17 | 24,725 | 45,381 | 150,505 | 248,265 |
| New York | 19 | 2,200 | 47,581 | 15,500 | 263,765 |
| Kentucky | 21 | 7,883 | 55,464 | 59,520 | 323,285 |
| Ohio | 21 | 15,827 | 71,291 | 123,431 | 446,716 |
| Montana | 81 | 1,881 | 73,172 | 55,577 | 502,293 |
| Illinois | 85 | 51 | 73,223 | 1,585 | 503,878 |
| Virginia | 91 | 258 | 73,481 | 8,544 | 512,422 |
| Colorado | 205 | 2,574 | 76,055 | 192,438 | 704,860 |
| Kansas | 207 | 10,544 | 86,599 | 798,563 | 1,503,423 |
| Arkansas | 215 | 1,402 | 88,001 | 109,923 | 1,613,346 |
| New Mexico | 239 | 13,598 | 101,599 | 1,185,999 | 2,799,345 |
| South Dakota | 252 | 10 | 101,609 | 919 | 2,800,264 |
| Oklahoma | 354 | 14,417 | 116,026 | 1,861,971 | 4,662,235 |
| Nebraska | 439 | 20 | 116,046 | 3,208 | 4,665,443 |
| Texas | 553 | 35,650 | 151,696 | 7,201,973 | 11,867,416 |
| California | 612 | 1,117 | 152,813 | 249,459 | 12,116,875 |
| Utah | 891 | 182 | 152,995 | 59,165 | 12,176,040 |
| Wyoming | 923 | 1,236 | 154,231 | 416,246 | 12,592,286 |
| Arizona | 953 | 1 | 154,232 | 348 | 12,592,634 |
| Alabama | 957 | 250 | 154,482 | 87,312 | 12,679,946 |
| Mississippi | 1,155 | 433 | 154,915 | 182,494 | 12,862,440 |
| Michigan | 1,185 | 376 | 155,291 | 162,691 | 13,025,131 |
| Louisiana | 1,381 | 14,464 | 169,755 | 7,292,278 | 20,317,409 |
| North Dakota | 1,595 | 78 | 169,833 | 45,400 | 20,362,809 |
| Alaska | 10,641 | 58 | 169,891 | 225,280 | 20,588,089 |

[a] Net wellhead production, dry gas and condensate wells only.

Source: Gas Facts, AGA, 1981, Tables 16 and 25.

In 1979, the petroleum industry spent approximately $16.1 billion for drilling and equipping onshore and offshore oil and gas wells and dry holes. The average depth of gas wells in 1979 was 5,493 feet. The average cost per foot was $80.66, up from $68.37 in 1978. Table 9-2 shows that the industry spent $12.8 billion for drilling and equipping 47,263 onshore oil and gas wells and dry holes.

The average depth of onshore gas wells in 1979 was 5,337 feet. In 1979, the average cost of drilling an onshore gas well was $367,710. Costs for drilling and equipping onshore gas wells through the "Christmas tree" (wellhead) are analyzed in Table 9-3 for selected depth intervals. Average costs per well by depth interval from the Joint Association Survey data are allocated a share of dry hole drilling costs and escalated 17.2 percent to obtain an estimated cost of drilling and equipping a producing gas well in 1980 dollars. At 2,000, 4,000, 8,000 and 12,000 feet, the estimated drilling costs per well are $89,000, $211,000, $949,000 and $1,875,000, respectively.[3]

Drilling costs are combined with Department of Energy estimates of additional field equipment costs and annual operating and maintenance costs. Total unit wellhead production costs are shown in Table 9-4. Field equipment costs and operating and maintenance costs are variable with gas flow rate. As shown in Table 9-4, field equipment costs (including dehydrators, field gathering pipelines, etc.) between the "Christmas tree" and point of transfer typically average $15,000 for a gas well producing 50 Mcfd and $64,000 for a well producing 5,000 Mcfd. Annual operating and maintenance costs average $6,000 per year for a gas well producing 50 Mcfd and $29,000 for a gas well producing 5,000 Mcfd. These costs are expressed in 1980 dollars.[4,5]

Estimated average unit total costs (1980 $/Mcf) for new wells at the selected depths and flow rates are shown in Table 9-4. These costs are based on the following assumptions: 20 percent royalty payments, 15 percent depletion allowance, 6 percent annual production decline rate, 6 percent annual operating cost decline rate, 20-year well life, 8 percent inflation, 10 percent nominal industry weighted cost of capital, 47 percent marginal corporate income tax rate, 10 percent investment tax credit rate,

Table 9-2. ESTIMATED COSTS OF DRILLING AND EQUIPMENT ONSHORE* WELLS, BY DEPTH INTERVALS - 1979

|  | DEPTH INTERVAL (FEET) | TOTAL NUMBER OF WELLS, FOOTAGE DRILLED, AND COSTS |  |  |  | AVERAGE DEPTH AND COST PER WELL |  |  |  |
|---|---|---|---|---|---|---|---|---|---|
|  |  | OIL WELLS | GAS WELLS | DRY HOLES | TOTAL | OIL | GAS | DRY | TOTAL |
| WELLS | 0 - 1,249 | 3,124 | 812 | 1,405 | 5,341 |  |  |  |  |
| FOOTAGE |  | 2,593,461 | 725,740 | 1,056,381 | 4,375,582 | 830 | 893 | 751 | 819 |
| COSTS $ |  | 86,386,485 | 22,857,939 | 28,102,986 | 137,347,410 | 27,652 | 28,150 | 20,002 | 25,715 |
| WELLS | 1,250 - 2,499 | 3,679 | 2,248 | 2,128 | 8,055 |  |  |  |  |
| FOOTAGE |  | 6,742,062 | 4,465,634 | 4,026,641 | 15,234,337 | 1,832 | 1,986 | 1,892 | 1,891 |
| COSTS $ |  | 241,163,201 | 142,528,882 | 95,923,692 | 479,615,775 | 65,551 | 63,402 | 45,076 | 59,542 |
| WELLS | 2,500 - 3,749 | 3,749 | 2,602 | 2,717 | 9,068 |  |  |  |  |
| FOOTAGE |  | 11,686,331 | 8,271,775 | 8,527,705 | 28,485,811 | 3,117 | 3,179 | 3,138 | 3,141 |
| COSTS $ |  | 401,962,239 | 273,680,295 | 193,938,405 | 869,580,939 | 107,218 | 105,180 | 71,379 | 95,895 |
| WELLS | 3,750 - 4,999 | 2,545 | 2,220 | 2,518 | 7,283 |  |  |  |  |
| FOOTAGE |  | 11,174,192 | 9,607,559 | 11,011,375 | 31,793,126 | 4,390 | 4,327 | 4,373 | 4,365 |
| COSTS $ |  | 410,452,099 | 314,233,848 | 257,064,007 | 981,749,954 | 161,277 | 141,546 | 102,090 | 134,800 |
| WELLS | 5,000 - 7,499 | 2,924 | 2,638 | 2,888 | 8,450 |  |  |  |  |
| FOOTAGE |  | 17,753,897 | 16,007,786 | 17,562,450 | 51,324,133 | 6,071 | 6,068 | 6,081 | 6,073 |
| COSTS $ |  | 806,370,701 | 685,068,394 | 514,402,953 | 2,005,842,048 | 275,776 | 259,692 | 178,117 | 237,377 |
| WELLS | 7,500 - 9,999 | 2,121 | 1,454 | 1,481 | 5,056 |  |  |  |  |
| FOOTAGE |  | 18,180,438 | 12,713,368 | 12,753,621 | 43,647,427 | 8,571 | 8,743 | 8,611 | 8,632 |
| COSTS $ |  | 1,085,358,592 | 839,442,485 | 580,625,002 | 2,505,426,079 | 511,720 | 577,333 | 392,049 | 495,535 |
| WELLS | 10,000 - 12,499 | 529 | 966 | 916 | 2,411 |  |  |  |  |
| FOOTAGE |  | 5,797,665 | 10,692,591 | 10,123,172 | 26,613,428 | 10,959 | 11,068 | 11,051 | 11,038 |
| COSTS $ |  | 549,408,187 | 1,028,150,996 | 719,506,625 | 2,297,065,808 | 1,038,578 | 1,064,338 | 785,487 | 952,744 |
| WELLS | 12,500 - 14,999 | 182 | 423 | 435 | 1,040 |  |  |  |  |
| FOOTAGE |  | 2,409,968 | 5,735,964 | 5,875,332 | 14,021,264 | 13,241 | 13,560 | 13,506 | 13,481 |
| COSTS $ |  | 285,191,467 | 746,345,379 | 598,673,500 | 1,630,210,346 | 1,566,986 | 1,764,409 | 1,376,260 | 1,567,509 |
| WELLS | 15,000 - 17,499 | 36 | 181 | 156 | 373 |  |  |  |  |
| FOOTAGE |  | 577,943 | 2,921,108 | 2,512,462 | 6,011,513 | 16,053 | 16,138 | 16,105 | 16,116 |
| COSTS $ |  | 95,730,408 | 520,415,261 | 363,138,240 | 979,283,959 | 2,659,178 | 2,875,222 | 2,327,809 | 2,625,426 |
| WELLS | 17,500 - 19,999 | 1 | 62 | 72 | 135 |  |  |  |  |
| FOOTAGE |  | 18,500 | 1,149,528 | 1,335,766 | 2,503,794 | 18,500 | 18,540 | 18,552 | 18,546 |
| COSTS $ |  | 2,625,523 | 268,628,618 | 294,558,344 | 565,812,505 | 2,625,523 | 4,332,719 | 4,091,088 | 4,191,203 |
| WELLS | 20,000 & OVER | 4 | 20 | 27 | 51 |  |  |  |  |
| FOOTAGE |  | 86,526 | 442,077 | 580,623 | 1,109,226 | 21,631 | 22,103 | 21,504 | 21,749 |
| COSTS $ |  | 19,108,100 | 169,068,312 | 231,787,935 | 419,964,347 | 4,777,025 | 8,453,415 | 8,584,738 | 8,234,595 |
| TOTAL WELLS |  | 18,894 | 13,626 | 14,743 | 47,263 |  |  |  |  |
| TOTAL FOOTAGE |  | 77,020,983 | 72,733,130 | 75,365,528 | 225,119,641 | 4,076 | 5,337 | 5,111 | 4,763 |
| TOTAL COSTS $ |  | 3,983,757,002 | 5,010,420,429 | 3,877,721,739 | 12,871,899,170 | 210,847 | 367,710 | 263,021 | 272,346 |

*Includes the following states and areas which are not shown separately: Arizona, Idaho, Missouri, Nevada, Oregon, South Dakota, Tennessee, and Virginia Note: Includes Alaska-Onshore

Source: API, 1979 Joint Association Survey on Drilling Costs, February, 1981 ed. Table 4

Table 9-3. ESTIMATED COST OF DRILLING AND EQUIPPING ONSHORE NATURAL GAS WELLS, BY SELECTED WELL DEPTHS, 1980

| Item | Unit | Well depth (feet) 2,000 | 4,000 | 8,000 | 12,000 |
|---|---|---|---|---|---|
| Cost per onshore gas well drilled | 1979$ | 63,000 | 142,000 | 578,000 | 1,064,000 |
| Dry hole factors [a] | | 1.2 | 1.3 | 1.4 | 1.5 |
| Cost per producing gas well drilled | 1979$ | 76,000 | 180,000 | 810,000 | 1,600,000 |
| Escalation factor [b] | | 1.172 | 1.172 | 1.172 | 1.172 |
| Cost per producing gas well drilled | 1980$ | 89,000 | 211,000 | 949,000 | 1,875,000 |

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[a] Dry hole cost factor equals the ratio of total cost divided by total costs for gas and oil wells, by depth interval

[b] American Petroleum Institute, 1979 Joint Association Survey on Drilling Costs, February 1981, edition. Table 3 page 9.

Table 9-4. ESTIMATED ONSHORE NATURAL GAS PRODUCTION COSTS FOR SELECTED WELL DEPTHS AND BASE YEAR FLOW RATES, 1980

| Flow rate and cost | | | | Well depth (feet) | | | |
|---|---|---|---|---|---|---|---|
| | | | | 2,000 | 4,000 | 8,000 | 12,000 |
| Drilling and equipment cost per productive gas well[a] (1980 $) | | | | 89,000 | 211,000 | 949,000 | 1,875,000 |
| Flow rate (Mcfd) | Field equipment cost[b c] (1980$) | Annual O&M cost[b] (1980 $/yr) | Unit variable cost[d] (1980$/Mcf) | Unit total cost<br>1980 $/Mcf | | | |
| 50 | 15,000 | 6,000 ($.36/Mcf) | .93 | 4.31 | 8.95 | 36.99 | 72.18 |
| 250 | 26,000 | 12,000 ($.14 Mcf) | .88 | 1.02 | 1.95 | 7.55 | 14.59 |
| 500 | 35,000 | 15,000 ($.09/Mcf) | .47 | .56 | 1.02 | 3.83 | 7.35 |
| 1,000 | 43,000 | 20,000 ($.06/Mcf) | .25 | .31 | .54 | 1.94 | 3.70 |
| 5,000 | 64,000 | 29,000 ($.02/Mcf) | .06 | .08 | .12 | .40 | .75 |
| 10,000 | 87,000 | 44,000 ($.01/Mcf) | .03 | .05 | .07 | .21 | .39 |

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[a] Source is Table 9-3.

[b] Source: "Aggregate Average Annual Gas Well Operating Costs vs Production Capacity 1981" DOE, Dallas, TX, (1981 figures deflated to 1980 by 9%) (See Reference 5).

[c] Includes field equipment between "Christmas tree" and point of transfer. Does not include sour gas sweetening or sulfur recovery equipment.

[d] Unit variable costs include field equipment costs and operating and maintenance costs.

and accelerated cost recovery system (over ten years).[13, 25, 26] The equation for estimating the average unit costs in Table 9-4 is shown in Appendix F.

The unit costs displayed in Table 9-4 illustrate the variability of new gas well production costs. The magnitude of the costs are indicative of average production costs under the assumed conditions. Variation of those conditions and individual well circumstances naturally affect actual costs.

The total unit costs of natural gas production from individual producing new wells range over at least four orders of magnitude (e.g., from $.05 per Mcf to over $70.00 per Mcf). Although the average costs of new natural gas for an individual producer, gas field, or composite for the industry is indeterminant from the data presented above due to a lack of new well characteristics (flow rates, decline rates, and depths), Table 9-4 does indicate several important characteristics of the industry. First, new gas production requires large front-end investments for exploration and development drilling. Second, there is a significant risk that any individual well will be dry or too small to recover total costs. Third, operating costs are relatively small, hence once drilling costs are sunk, it is often economical for the operator to operate the well to cover short run variable costs while making some contribution, however small, to capital recovery. Consequently, some wells produce at a loss against total costs. Economic profits on individual wells or gas fields are inversely related to well depth and positively related to well flow rate, all other things being equal.

### 9.1.1.3 Natural Gas Sulfur Recovery Facilities

In 1980, there were 31 companies that owned a total of 89 onshore natural gas sulfur recovery facilities. These facilities had a total capacity of approximately 10,600 megagrams of recovered sulfur per day. Table 9-5 presents these 31 companies, their sulfur recovery facilities, and their combined daily capacity.[6] Table 9-5 indicates that no single company operator controls the industry. However, a major fraction of the combined sulfur recovery capacity is controlled by a relatively small number of companies. The five largest operators have a combined capacity of 8,238 megagrams of recovered sulfur per day, which is 77.4 percent of

Table 9-5 ONSHORE NATURAL GAS SULFUR RECOVERY FACILITY OPERATORS, 1979

| Number | Company/operator | Number of operating facilities | Daily sulfur intake capacity (combined), Mg/d |
|---|---|---|---|
| 1 | Shell Oil Company | 7 | 2,207 |
| 2 | Exxon Company USA | 7 | 1,974 |
| 3 | Amoco Production Company | 10 | 1,962 |
| 4 | Pursue Gas Processing and Petroleum Company (Mid-American Oil and Gas and Pursue Energy Corporation) | 1 | 1,095 |
| 5 | Chevron U.S.A., Inc. (Standard Oil Company of California) | 1 | 1,000 |
| 6 | Getty Oil Company | 3 | 354 |
| 7 | Warren Petroleum Company (Gulf Oil Corporation) | 6 | 281 |
| 8 | Phillips Petroleum Company | 5 | 279 |
| 9 | El Paso Natural Gas Company | 5 | 276 |
| 10 | Cities Service Company | 4 | 271 |
| 11 | Aminoil U.S.A., Inc. (R.J. Reynolds Industries, Inc.) | 2 | 230 |
| 12 | Texaco, Inc. | 9 | 90.3 |
| 13 | Husky Oil Company | 3 | 85 |
| 14 | Trans-Jeff Chemical Corporation | 1 | 80 |
| 15 | Marathon Oil Company | 2 | 66 |
| 16 | Colorado Interstate Gas Company (Coastal States Gas Corporation) | 1 | 66 |
| 17 | Atlantic Richfield Company | 2 | 49 |
| 18 | Natural Gas Pipeline Company of America | 4 | 47 |

(continued)

Table 9-5. Continued

| Number | Company/operator | Number of operating facilities | Daily sulfur intake capacity (combined), Mg/d |
|---|---|---|---|
| 19 | Tonkawa Gas Processing Company (Texas Oil and Gas Corporation) | 2 | 35 |
| 20 | Intratex Gas Company (Houston Natural Gas Company) | 1 | 33 |
| 21 | Diamond Shamrock Corporation | 1 | 30 |
| 22 | Sinclair Oil Corporation (Little America Refining Company) | 1 | 25 |
| 23 | MGF Oil Corporation | 1 | 22 |
| 24 | Union Oil Company of California | 1 | 18 |
| 25 | Lone Star Gas Company | 1 | 15 |
| 26 | Northern Natural Gas Company | 1 | 14 |
| 27 | Suburban Propane Gas Corporation | 1 | 14 |
| 28 | Pioneer Gas Products Company (Pioneer Corporation) | 2 | 9 |
| 29 | Westland Oil Development Corporation | 1 | 5.5 |
| 30 | Gulf Energy and Development Corporation | 1 | 5 |
| 31 | Dorchester Gas Corporation | 2 | 3 |
|  | TOTAL | 89 | 10,640.8 |

Source: Chemical Economics Handbook - Stanford Research Institute (SRI) International, December, 1979.

the total capacity of the industry. In addition, the 15 largest operators have a combined capacity of 10,250 megagrams per day or 96.3 percent of the total industry capacity. The remaining 16 operators have a combined capacity of 3.7 percent of the total. This distribution implies that small producers tend to leave processing activities to the larger firms. Some of these smaller operators are gas transmission (pipeline) companies engaged in the production of oil and gas. However, most of these sulfur recovery facility operators are diversified oil and gas producing companies and are ranked among the largest 100 industrial corporations listed by Fortune Magazine.

Appendix G (Table G-8) presents the American Petroleum Institute's July 1982 survey on the onshore natural gas processing facilities. Data on cpacity, throughout, as well as $H_2S$ and $CO_2$ mole percent are listed for each gas stream. Table 9-6 summarizes the observed frequency of sweetening facilities by $H_2S$ (percent) in sour natural gas and by facility capacity (MG/day). The two-way frequency distribution is for streams with sweetening only. (The underlying data are presented in Table G-2). Table 9-6 indicates that small capacity facilities tend to be associated with very lean $H_2S$ percentages in the sour natural gas. Table 9-7 presents the observed frequency of sulfur recovery facilities by $H_2S$ (percent) in sour natural gas and by facility capacity (Mg/day). This two-way frequency distribution is for streams with sulfur recovery only. (The underlying data are presented in Table G-3). Table 9-7 indicates that facilities with large sulfur recovery capacities tend to be associated with high $H_2S$ percentages in the sour natural gas. Table 9-6 and 9-7 also demonstrate that the $H_2S$ concentrations range widely from less than 0.1 to above 15 mole percent. A higher $H_2S$ percentage yields more sulfur and less sweet gas (residue gas) and vice versa.

In 1980, 1,707,000 megagrams of sulfur were recovered from onshore natural gas, which is equivalent to 4,877 megagrams per day (1 year = 350 days). This represents 72.5 percent industry capacity utilization. The percentage of capacity utilization in the industry for the years 1951 through 1980, the number of new facilities added each year, and their capacities are presented in Table 9-8.

## Table 9-6 OBSERVED FREQUENCY OF SWEETENING FACILITIES BY $H_2S$ PERCENTAGE IN SOUR NATURAL GAS AND FACILITY CAPACITY, JULY 1982

| Facility capacity, Mg/d | $H_2S$ percentage in sour natural gas | | | | | | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | <0.1 | 0.1 | 0.2 | 0.3 | 0.4 | 0.5 | 0.6 | 0.7 | 0.8 | 0.9 | 1 | 2 | 3 | 4 | 5 | >5 | Total |
| <0.1 | 51 | | 1 | 1 | | | | | | | | | | | | | 53 |
| 0.1 | 2 | | 1 | | | | | | | | | | | | | | 3 |
| 0.2 | 11 | | 1 | | | | 1 | | | | | | | | | | 13 |
| 0.3 | 5 | | | | 1 | 1 | | | | | | | | | | | 7 |
| 0.4 | 4 | 2 | | | | 2 | | 1 | | | | | | | | | 9 |
| 0.5 | 3 | | 1 | | 3 | | | | | | | | | | | | 7 |
| 0.6 | 3 | | 1 | | | 2 | | | | | | | | | | | 6 |
| 0.7 | 3 | | | 1 | | | | | | | | | | | | | 4 |
| 0.8 | 1 | | | | 1 | | | | | | | | | | | | 2 |
| 0.0 | 1 | 2 | | | | | | | | | | | | | | | 3 |
| 1 | 2 | 4 | 3 | 2 | | 1 | | | 9 | | | 1 | | | | 1 | 23 |
| 2 | 4 | 1 | 1 | 1 | 5 | 2 | | | | | 2 | 1 | | | | | 17 |
| 3 | 4 | | 2 | 3 | | | 1 | | 5 | | | 1 | 1 | | | | 17 |
| 4 | 1 | | 1 | 3 | | 2 | | 1 | | | 2 | 2 | 1 | | | | 13 |
| 5 | | | | | | | | | | | | | | | | | 0 |
| >5 | 2 | | 7 | 1 | 3 | 2 | 1 | 1 | 1 | | 6 | 6 | 2 | | 2 | 2 | 36 |
| Total | 97 | 9 | 19 | 12 | 13 | 12 | 3 | 3 | 15 | 0 | 10 | 11 | 4 | 0 | 2 | 3 | 213 |

**Source:** Appendix G, Table G-2.

## Table 9-7. OBSERVED FREQUENCY OF SULFUR RECOVERY FACILITIES BY $H_2S$ PERCENTAGE IN SOUR NATURAL GAS AND FACILITY CAPACITY, JULY 1982

| Facility capacity, Mg/d | $H_2S$ percentage in sour natural gas | | | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | ≤0.5 | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11-15 | >15 | Total |
| <10 | 3 | 1 | 2 | | 1 | 2 | | | 1 | 1 | | | | 11 |
| 10-49 | 5 | 7 | 5 | 3 | 4 | 3 | 2 | | 1 | | | 3 | | 33 |
| 50-99 | | | | | 4 | | | | | | 3 | | 2 | 9 |
| 100-199 | | | | | | | 2 | | 1 | | | | 1 | 4 |
| >200 | | | | | | | | | | | | 1 | 3 | 4 |
| Total | 8 | 8 | 7 | 3 | 9 | 5 | 4 | | 3 | 1 | 3 | 4 | 6 | 61 |

**Source:** Appendix G, Table G-3.

### Table 9-8. ONSHORE NATURAL GAS SULFUR RECOVERY FACILITIES AND SULFUR INTAKE CAPACITY UTILIZATION, 1950-1982

| Year | Cumulative Number of operating facilities | Newly constructed facilities | | Cumulative capacity, Mg/d | Sulfur recovered, $10^3$ Mg | Percent capacity utilization[a] |
|---|---|---|---|---|---|---|
| | | Number | New capacity added, Mg/d | | | |
| Before 1950 | 4 | 4 | 254 | 254 | N.A.[b] | –[c] |
| 1951-1955 | 9 | 5 | 323 | 577 | 178.3<br>(1955) | 88.3 |
| 1956-1960 | 16 | 7 | 643 | 1,220 | 330<br>(1960) | 77.3 |
| 1961-1965 | 30 | 14 | 923 | 2,143 | 522<br>(1965) | 69.6 |
| 1966-1970 | 45 | 15 | 1,404 | 3,547 | 627<br>(1970) | 50.5 |
| 1971 | 48 | 3 | 22 | 3,569 | 638 | 51.1 |
| 1972 | 53 | 5 | 1,420 | 4,989 | 819 | 46.9 |
| 1973 | 58 | 5 | 115 | 5,104 | 1,046 | 58.6 |
| 1974 | 67 | 9 | 1,210.4 | 6,314.4 | 1,219 | 55.2 |
| 1975 | 70 | 3 | 85.5 | 6,399.9 | 1,364 | 60.9 |
| 1976 | 73 | 3 | 40.5 | 6,440.4 | 1,298 | 57.6 |
| 1977 | 74 | 1 | 25 | 6,465.4 | 1,426 | 63.0 |
| 1978 | 78 | 4 | 95.7 | 6,561.1 | 1,753 | 76.3 |
| 1979 | 83 | 5 | 83.7 | 6,644.8 | 1,760 | 75.7 |
| 1980 | 85 | 2 | 81 | 6,725.8 | 1,707 | 72.6 |
| 1981 | 87 | 2 | 1,765 | 8,490.8 | N.A. | – |
| 1982 | 89 | 2 | 2,150 | 10,640.8 | N.A. | – |

[a] Derived values based on 350 days/year operating schedule.

[b] Not available.

[c] Not applicable.

The natural gas sulfur recovery industry has added 1 to 9 new facilities each year since 1971. The addition of new operating facilities is presented in Figure 9-1 for each year from 1971 through 1982 and for each 5-year period from 1950 through 1970. Similarly, the total new sulfur recovery capacity added for each 5-year period from 1950 through 1970 and for each year from 1971 through 1982 is shown in Figure 9-2.

Not all capacity additions mean expanded output. The capacity of a sulfur recovery facility is generally designed to accommodate the highest potential sulfur input to the facility. Facilities are designed this way in order to insure that the primary operation at the site will never be shut down because of failure of the sulfur recovery operation.

The size distribution for seven categories of existing sulfur recovery facilities is presented in Table 9-9. By 1982, sixty-two facilities, representing 69.7 percent of the total, will fall in the smallest two categories (less than 10 and 10 to 49 Mg/day), and the average capacity of these facilities will be approximately 10.2 megagrams per day. The fifth and sixth categories (200 to 499 and 500 to 999 Mg/day) include seven facilities (7.9 percent of total) with an average capacity of approximately 564 megagrams per day. The last category (1,000 and above Mg/day) includes four facilities (4.5 percent of total) with an average capacity of approximately 1,016 megagrams per day. Table 9-9 also indicates that the capacity distribution at the end of 1972 is about the same as that at the end of 1982. Therefore, the capacity distribution estimated at the end of 1982 is projected to remain unchanged for the period from 1983 through 1987 and is used as the basis for further analyses.

Sulfur recovery facilities are not uniformly located across the United States. Typically, the facility is operated at the field where the sour natural gas is produced. Forty-nine facilities are located in Texas and 10 facilities in New Mexico. The Mississippi-Alabama-Florida region contains 13 facilities and Wyoming has 8 facilities.

Employment in natural gas sulfur recovery is not labor intensive. The operations are automated and, therefore, are controlled by instrumentation. Approximately 10 persons are employed in labor, supervision, maintenance/repair and administration in a typical onshore natural gas sulfur recovery facility of 100 megagrams sulfur intake per day. Installation of

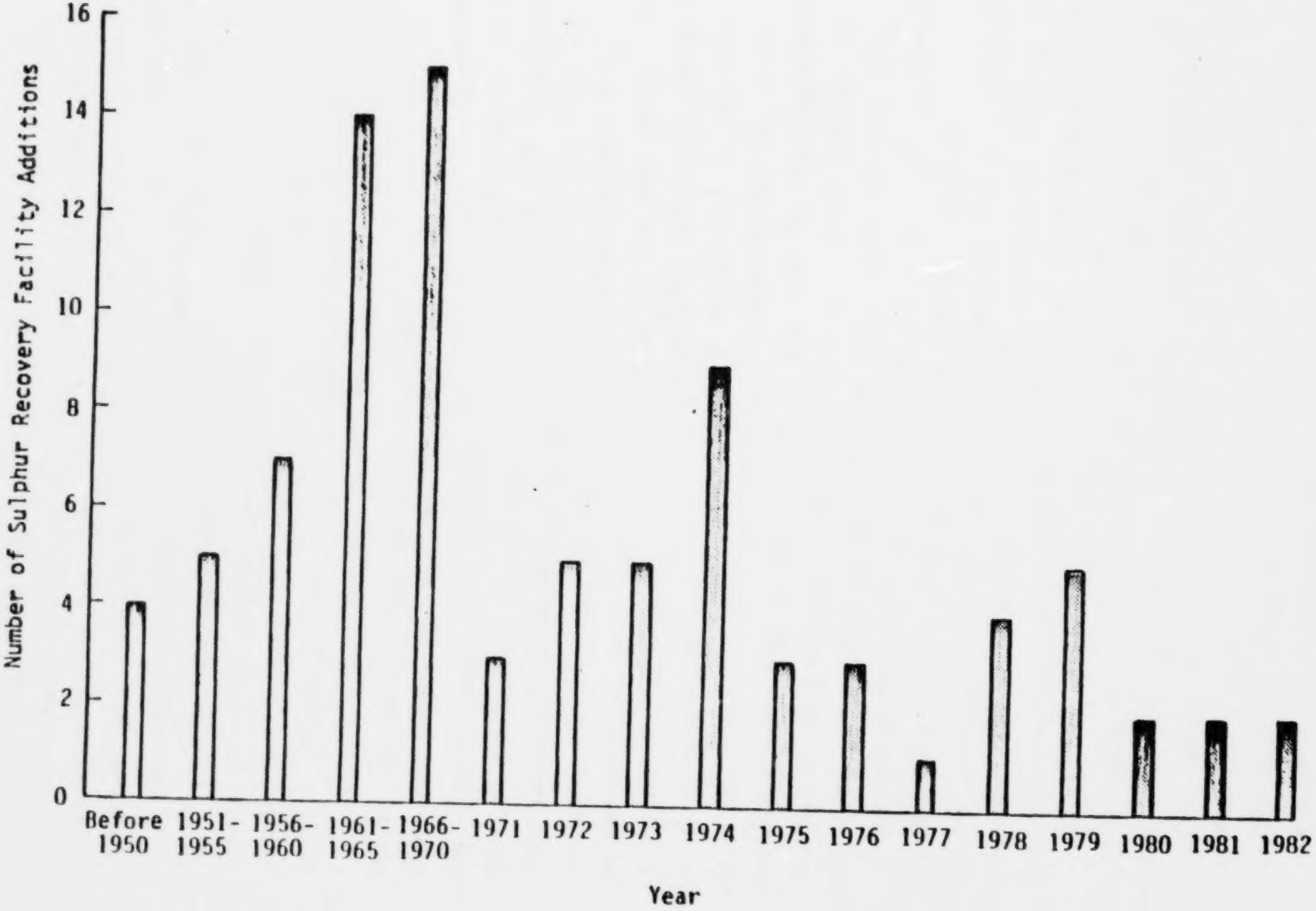

Figure 9-1. Sulfur recovery facility additions, 1950-1982.

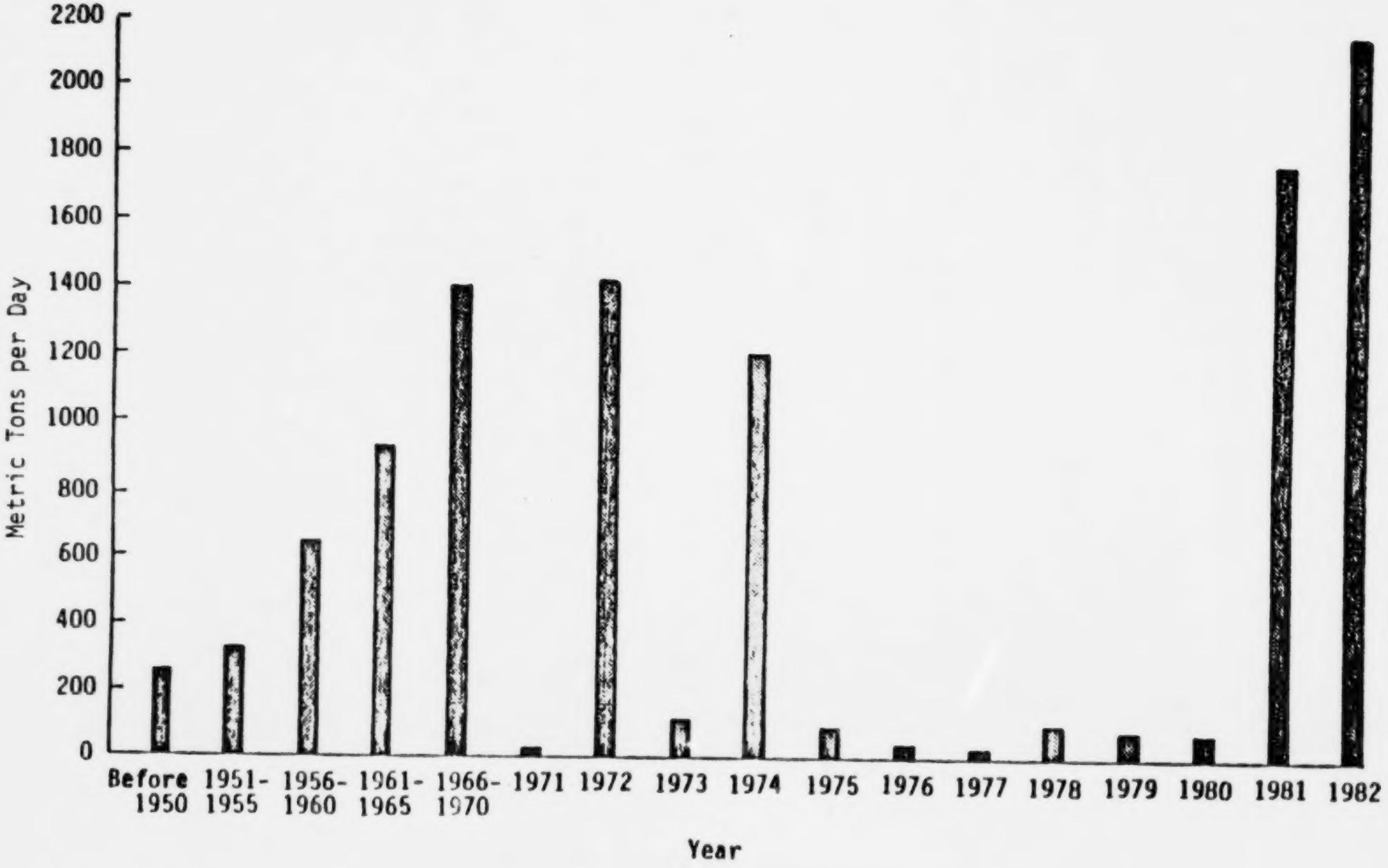

Figure 9-2. Sulfur recovery capacity additions, 1950-1982.

## Table 9-9. SULFUR INTAKE CAPACITY DISTRIBUTION OF ONSHORE NATURAL GAS SULFUR RECOVERY FACILITIES, 1950-1982

| Year | Cumulative number of facilities | | | | | | |
|---|---|---|---|---|---|---|---|
| | Sulfur intake size range | | | | | | |
| | < than 10, Mg/d | 10-49, Mg/d | 50-99, Mg/d | 100-199, Mg/d | 200-499, Mg/d | 500-999, Mg/d | 1,000 and >, Mg/d |
| Before 1950 | – | 2 | 1 | 1 | – | – | – |
| 1951-1955 | – | 6 | 1 | 1 | 1 | – | – |
| 1956-1960 | 1 | 7 | 4 | 2 | 2 | – | – |
| 1961-1965 | 3 | 15 | 6 | 3 | 3 | – | – |
| 1966-1970 | 7 | 23 | 6 | 4 | 4 | 1 | – |
| 1971 | 9 | 24 | 6 | 4 | 4 | 1 | – |
| 1972 | 11 | 25 | 6 | 5 | 4 | 1 | 1 |
| 1973 | 12 | 28 | 6 | 5 | 4 | 1 | 1 |
| 1974 | 15 | 31 | 7 | 6 | 4 | 1 | 1 |
| 1975 | 16 | 33 | 8 | 6 | 4 | 2 | 1 |
| 1976 | 17 | 35 | 8 | 6 | 4 | 2 | 1 |
| 1977 | 17 | 36 | 8 | 6 | 4 | 2 | 1 |
| 1978 | 18 | 38 | 9 | 6 | 4 | 2 | 1 |
| 1979 | 19 | 42 | 9 | 6 | 4 | 2 | 1 |
| 1980 | 19 | 43 | 10 | 6 | 4 | 2 | 1 |
| 1981 | 19 | 43 | 10 | 6 | 4 | 3 | 1 |
| 1982 | 19 | 43 | 10 | 6 | 4 | 3 | 4 |
| Percent distribution at the end of 1972 | 20.8 | 47.2 | 11.3 | 9.4 | 7.5 | 1.9 | 1.9 |
| Percent distribution at the end of 1982 | 21.4 | 48.3 | 11.2 | 6.7 | 4.5 | 3.4 | 4.5 |

Source: Chemical Economics Handbook - Stanford Research Institute (SRI) International, December 1979.

additional control equipment to enhance sulfur recovery and thereby reduce sulfur compounds emissions should not significantly increase the number of employees per facility.

9.1.1.4 Natural Gas Market. Although the natural gas component of total domestic energy production has decreased from 40 percent in 1973 to 34 percent in 1980 as indicated in Table 9-10, the natural gas production industry is expected to continue to supply a significant fraction of total domestic energy demand. Exploration and production activities for natural gas are anticipated to continue to increase as a result of phased natural gas price deregulation and expected price increases. The U.S. supply of natural gas from domestic production is price inelastic, according to Neri (1977) who reports that the 1980 implied price elasticities for domestic natural gas production are 0.06 and 0.24 for the American Gas Association (AGA) TERA Model (1973) and the MacAvoy-Pindyck (M-P) models (1975), respectively. Both models are based on similar theoretical concepts with respect to exploration and discovery processes but differ in estimation of the processes. The TERA model relies upon econometric and engineering elements while the M-P model is wholly econometric in approach. Both deal with the long-run dynamic behavior of drilling, new discoveries, reserve additions, and production.[7]

Domestic aggregate retail price elasticities of demand for solid fuels, natural gas, electricity and petroleum are shown in Table 9-11. These elasticities represent the change in final demand for each fuel with respect to a change in the price of all four aggregate fuel types. Therefore, the diagonal corresponding to direct price elasticity should have a negative sign. For example, the domestic retail price elasticity for natural gas is -.426, indicating a rather price inelastic aggregate retail demand. Electricity has the highest cross price elasticity with respect to natural gas with a value of .228, indicating that a one percent increase in the retail natural gas price causes a .228 percent increase in the aggregate quantity demanded of electricity. All of the cross price elasticities are positive, representing interfuel substitution.[8]

Table 9-10. PRODUCTION OF ENERGY BY TYPE, UNITED STATES (Quadrillion Btu)

| | Coal[a] | Crude oil[b] | NGPL[c] | Natural gas (dry) | Hydro-electric power[d] | Nuclear electric power | Other[e] | Total energy produced | % NG of |
|---|---|---|---|---|---|---|---|---|---|
| 1973 | 14.366 | 19.493 | 2.569 | 22.187 | 2.861 | 0.910 | 0.046 | 62.433 | 40 |
| 1974 | 14.468 | 18.575 | 2.471 | 21.210 | 3.177 | 1.272 | 0.056 | 61.229 | 39 |
| 1975 | 15.189 | 17.729 | 2.374 | 19.640 | 3.155 | 1.900 | 0.072 | 60.059 | 37 |
| 1976 | 15.853 | 17.262 | 2.327 | 19.480 | 2.976 | 2.111 | 0.081 | 60.091 | 36 |
| 1977 | 15.829 | 17.454 | 2.327 | 19.565 | 2.333 | 2.702 | 0.082 | 60.293 | 36 |
| 1978 | 15.037 | 18.434 | 2.245 | 19.485 | 2.958 | 2.977 | 0.068 | 61.204 | 36 |
| 1979 | 17.651 | 18.104 | 2.286 | 20.076 | 2.954 | 2.748 | 0.089 | 63.907 | 35 |
| 1980 | 18.877 | 18.250 | 2.263 | 19.754 | 2.913 | 2.704 | 0.114 | 64.876 | 34 |

Totals may not equal sum of components due to independent rounding.

[a] Includes bituminous coal, lignite and anthracite.
[b] Includes lease condensate.
[c] Natural gas plant liquids.
[d] Includes industrial and utility production of hydropower.
[e] Includes geothermal power and electricity produced from wood and waste.
R = Revised data
Source: U.S. Department of Energy, Energy Information Administration calculations. Monthly Energy Review, July 1981.

### Table 9-11. AGGREGATE RETAIL PRICE ELASTICITIES OF DEMAND, U.S. (Estimate for 1985)

| | Price elasticity of demand | | | |
|---|---|---|---|---|
| With respect to | Solid fuels | Natural gas | Electricity | Petroleum |
| Solid fuels | -.215 | .030 | .131 | .031 |
| Natural gas | .005 | -.426 | .228 | .062 |
| Electricity | .011 | .052 | -.376 | .111 |
| Petroleum | .002 | .013 | .077 | -.263 |

Source: The Global 2000 Report to the President, (Volume III: Documentation), A report prepared by the Council on Environmental Quality and the Department of State. April 1981. p. 301.

The U.S. is a net importer of natural gas and imports have remained fairly constant since 1973, ranging from 953 billion cubic feet in 1975 to 1,253 billion cubic feet in 1979. In 1980, the U.S. imported 994 billion cubic feet of natural gas primarily from Canada, Mexico, and Algeria. Exports of natural gas declined from 77 billion cubic feet in 1973 to 55 billion cubic feet in 1980. Exports are primarily to Japan and Mexico.

9.1.1.5 Sulfur Market. Over 90 percent of sulfur produced in the U.S. is consumed as feedstock in the manufacturing of sulfuric acid, which in turn is utilized in the fertilizer industry and other industries. Demand for sulfur is derived from the demand for sulfuric acid and is likely to be somewhat inelastic with respect to price but elastic with respect to industrial production and income in sulfur consuming industries.[10]

Domestic sulfur production from various sources for the period from 1950 through 1980 is shown in Figure 9-3. This figure indicates that Frasch mines are the main source of domestic sulfur. Over the last 30 years, Frasch production has increased at an uneven pace reaching a peak in 1974 at approximately 8 million metric tons. The Frasch mining segment of the sulfur industry is experiencing high energy costs and resource depletion. The share of total sulfur production that is recovered from petroleum refineries, natural gas sulfur recovery, and smelter gases has increased from 2 percent of the total in 1950 to over 38 percent in 1978. These trends are shown in Figure 9-4 and Table 9-12. Table 9-12 also shows that in 1978 natural gas sulfur recovery ranked as the fourth most important source of sulfur production following Frasch mines, recovery from petroleum refineries, and elemental imports.[9]

Recovered sulfur is expected to show continued steady growth because of environmental considerations. Sulfur recovery that is required by environmental regulations is effectively nondiscretionary, usually being the by-product of another industry. Consequently, the recovered elemental sulfur segment of total sulfur supply tends to be highly price inelastic. Mined sulfur and imports are likely to be more price elastic than recovered sulfur.

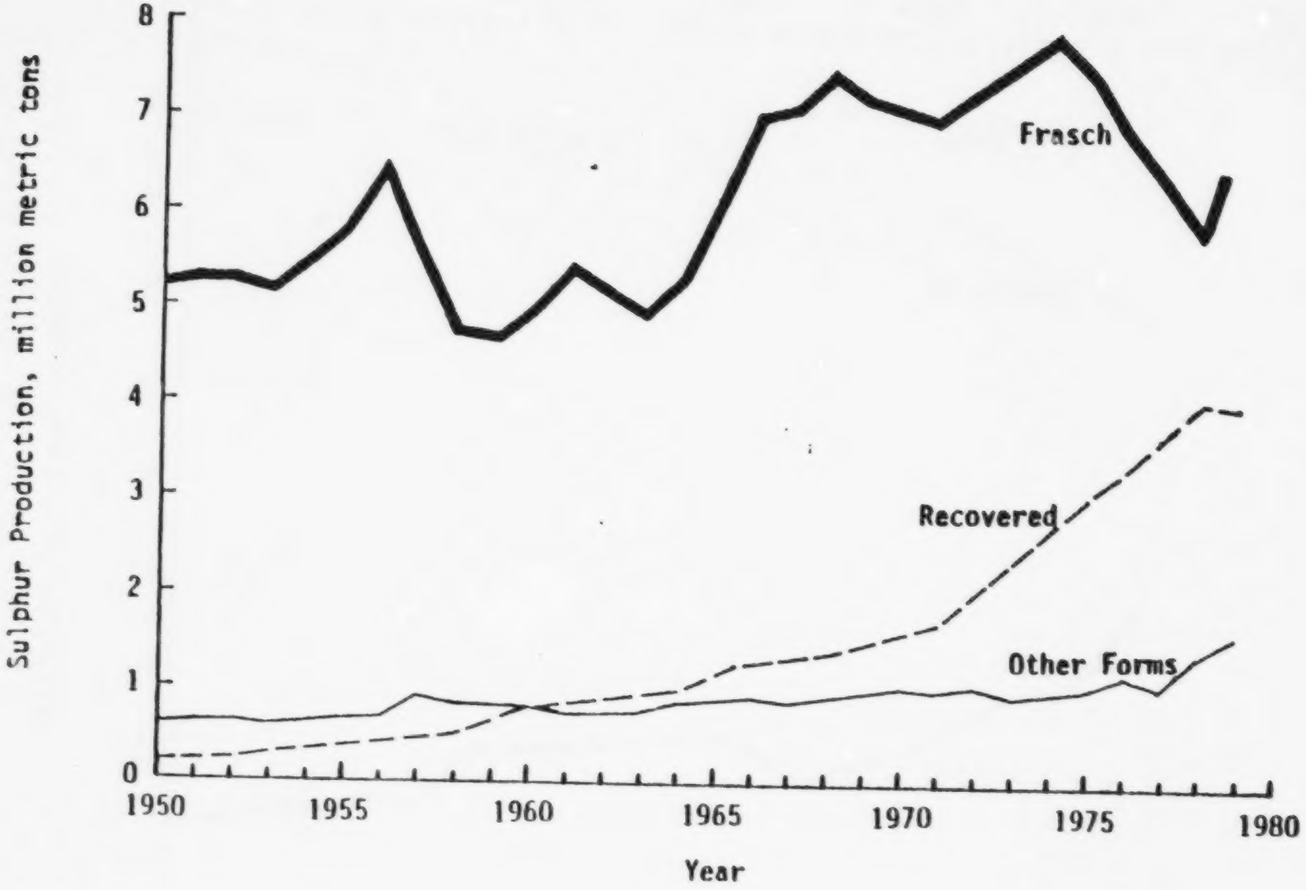

Figure 9-3. U.S. domestic sulfur production from various sources for the period 1950-1980.[7]

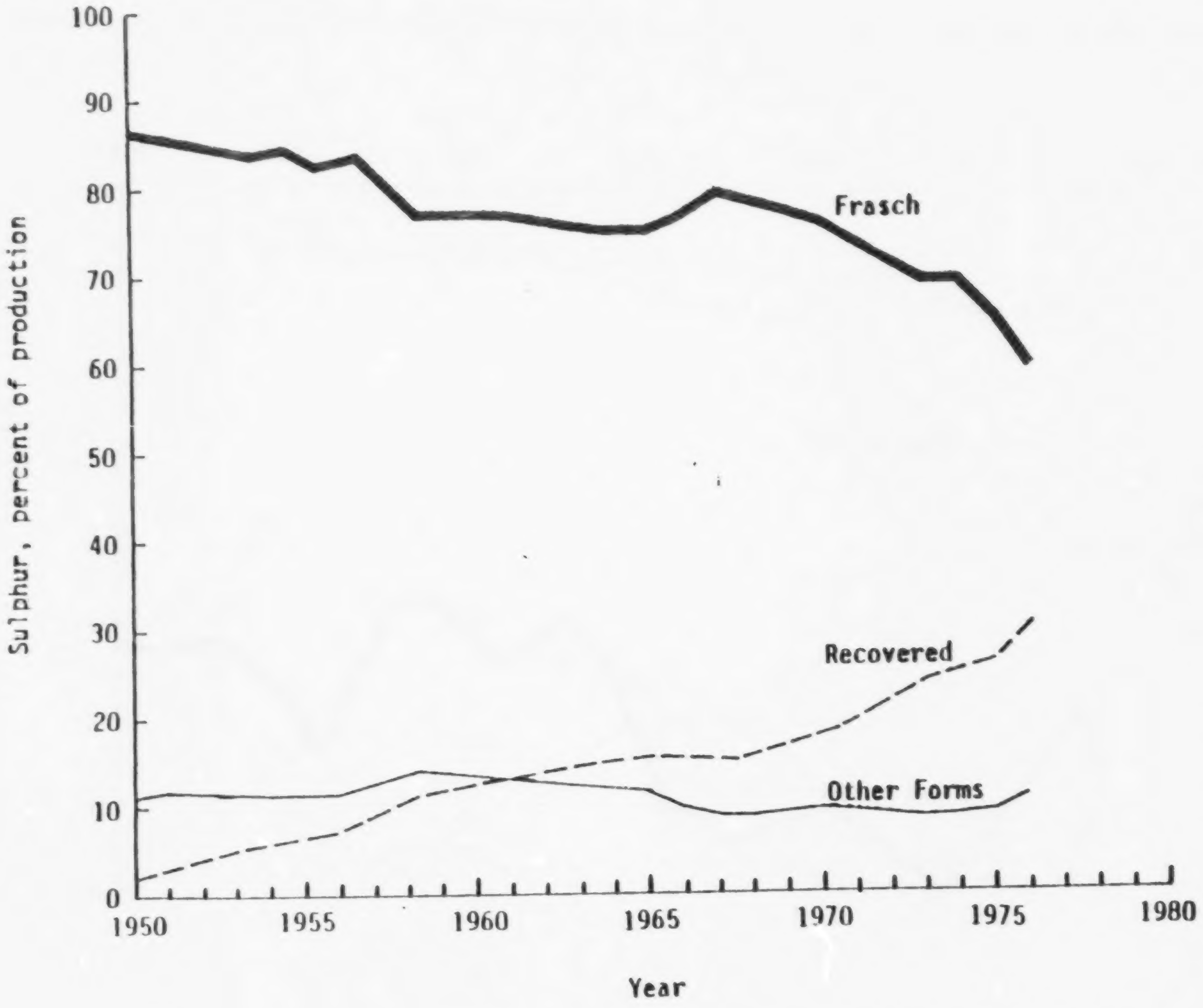

Figure 9-4. Trends in the production of sulfur in the United States.[9]

## Table 9-12. DOMESTIC SULFUR SUPPLY - 1978 STATISTICS

| Source | Percent of Total |
|---|---|
| (1) Frasch mining | 41.4 |
| (2) Recovered from petroleum refineries | 16.9 |
| (3) Recovered from natural gas | 12.9 |
| (4) Smelter gases | 8.1 |
| (5) Other[a] | 4.7 |
| (6) Elemental imports | 16.0 |

[a] Includes pyrites (imported and domestic), stocks, and various sulfur-containing compounds produced directly without any elemental sulfur requirements. Not included are imports of such sulfur-containing materials as $H_2SO_4$ and liquid $SO_2$.

Source: Chemical Economics Handbook - Stanford Research Institute (SRI) International, December 1979.

Exports of sulfur declined to 1.6 million megagrams in 1982, while imports remained steady at 2.5 million megagrams. Annual sulfuric acid exports are small, less than 100,000 megagrams, and are primarily to Canada and Mexico. Sulfuric acid imports are also small, under 500,000 megagrams per year, and are primarily from Canada.

### 9.1.2 The Natural Gas Sulfur Recovery Industry--Growth and Projections

This section discusses the historical production of natural gas, the price history of natural gas, the production history of liquid sulfur recovered from the natural gas production operations, and the price history of liquid sulfur. Natural gas production is projected for the years 1985, 1990 and 2000 and distributed to the categories of onshore, offshore, discoveries from existing fields, and new fields. This section also presents the new sulfur recovery capacity added each year for the period from 1950 through 1982 and the projected number, sizes and capacities of new facilities that are projected to be constructed and in operation during the period from 1983 through 1987.

9.1.2.1 Historical Data. Marketed production of natural gas increased from 5.42 trillion cubic feet in 1949 to a peak of 22.65 trillion cubic feet in 1973, an average of 6.0 percent annually. In 1974 and 1975, marketed production decreased 4.6 percent and 6.9 percent, respectively. After 1976, marketed production declined slightly to 19.7 trillion cubic feet in 1979.

Total gross withdrawals of natural gas from both gas wells and oil wells generally follow the same trend as marketed production. However, the volume of natural gas withdrawn from oil wells has remained relatively constant at about three to five trillion cubic feet per year from 1949 to the present. Table 9-13 presents total natural gas production distributed between onshore and offshore production for the years 1949 through 1979.[11] Onshore production declined from 99.1 percent of the total in 1954 to 72.4 percent of the total in 1979. Figure 9-5 shows natural gas gross withdrawals and marketed production from gas wells and oil wells from 1949 through 1979.[12] The difference between gross withdrawals and marketed production represents quantities from gas wells and oil wells that were either vented, flared or used for reservoir repressuring. In 1978, there

### Table 9-13. NATURAL GAS GROSS WITHDRAWALS AND MARKETED ONSHORE AND OFFSHORE PRODUCTION, 1949-1979

| | Production in Trillion Cubic Feet | | | | | | Percentage | |
|---|---|---|---|---|---|---|---|---|
| Year | From Gas Wells | From Oil Wells | Gross Withdrawals | Marketed Production[a] | Onshore Production | Offshore Production[c] | Onshore | Offshore |
| 1949 | 4.99 | 2.56 | 7.55 | 5.42 | NA | NA | NA | NA |
| 1950 | 5.60 | 2.88 | 8.48 | 6.28 | NA | NA | NA | NA |
| 1951 | 6.48 | 3.21 | 9.69 | 7.46 | NA | NA | NA | NA |
| 1952 | 6.84 | 3.43 | 10.27 | 8.01 | NA | NA | NA | NA |
| 1953 | 7.10 | 3.55 | 10.65 | 8.40 | NA | NA | NA | NA |
| 1954 | 7.47 | 3.52 | 10.98 | 8.74 | 8.66 | 0.08 | 99.1 | 0.9 |
| 1955 | 7.84 | 3.88 | 11.72 | 9.41 | 9.28 | 0.13 | 98.6 | 1.4 |
| 1956 | 8.31 | 4.07 | 12.37 | 10.08 | 9.94 | 0.14 | 98.6 | 1.4 |
| 1957 | 8.72 | 4.19 | 12.91 | 10.68 | 10.51 | 0.17 | 98.4 | 1.6 |
| 1958 | 9.15 | 3.99 | 13.15 | 11.03 | 10.77 | 0.26 | 97.6 | 2.4 |
| 1959 | 10.10 | 4.13 | 14.23 | 12.05 | 11.70 | 0.35 | 97.1 | 2.9 |
| 1960 | 10.85 | 4.23 | 15.09 | 12.77 | 12.33 | 0.44 | 96.6 | 3.4 |
| 1961 | 11.20 | 4.27 | 15.46 | 13.25 | 12.77 | 0.48 | 96.4 | 3.6 |
| 1962 | 11.70 | 4.34 | 16.04 | 13.88 | 13.24 | 0.64 | 95.4 | 4.6 |
| 1963 | 12.61 | 4.37 | 16.97 | 14.75 | 13.99 | 0.76 | 94.8 | 5.2 |
| 1964 | 13.11 | 4.43 | 17.54 | 15.55 | 14.70 | 0.85 | 94.5 | 5.5 |
| 1965 | 13.52 | 4.44 | 17.96 | 16.04 | 15.10 | 0.94 | 94.1 | 5.9 |
| 1966 | 13.89 | 5.14 | 19.03 | 17.21 | 15.84 | 1.37 | 92.0 | 8.0 |
| 1967 | 15.35 | 4.91 | 20.25 | 18.17 | 16.33 | 1.84 | 89.9 | 10.1 |
| 1968 | 16.54 | 4.79 | 21.33 | 19.32 | 17.00 | 2.32 | 88.0 | 12.0 |
| 1969 | 17.49 | 5.19 | 22.68 | 20.70 | 17.86 | 2.84 | 86.3 | 13.7 |
| 1970 | 18.59 | 5.19 | 23.79 | 21.92 | 18.70 | 3.22 | 85.3 | 14.7 |
| 1971 | 18.93 | 5.16 | 24.09 | 22.49 | 18.74 | 3.75 | 83.2 | 16.8 |
| 1972 | 19.04 | 4.97 | 24.02 | 22.53 | 18.77 | 3.76 | 83.3 | 16.7 |
| 1973 | 19.37 | 4.70 | 24.07 | 22.65 | 18.67 | 3.98 | 82.4 | 17.6 |
| 1974 | 18.67 | 4.18 | 22.85 | 21.60 | 17.37 | 4.23 | 80.4 | 19.6 |
| 1975 | 17.38 | 3.72 | 21.10 | 20.11 | 15.85 | 4.26 | 78.8 | 21.2 |
| 1976 | 17.19 | 3.75 | 20.94 | 19.95 | 15.65 | 4.30 | 78.4 | 21.6 |
| 1977 | 17.42 | 3.68 | 21.10 | 20.03 | 15.49 | 4.54 | 77.3 | 22.7 |
| 1978 | 17.39 | 3.91 | 21.31 | 19.97 | 14.87 | 5.10 | 74.5 | 25.5 |
| 1979[b] | 17.17 | 3.75 | 20.92 | 19.67 | 14.25 | 5.42 | 72.4 | 27.6 |

NA = Not Available.

[a] Marketed production is derived. It is gross withdrawals from producing reservoirs, less gas used for reservoir repressurizing and quantities vented and flared.

[b] Estimated, based on reported data through November.

[c] Data from U.S. Department of the Interior, Geological Survey - Conservation Division, Outer Continental Shelf Statistics.

Note: Sum of components may not equal total due to independent rounding. Beginning with 1965 data, all volumes are shown on a pressure base of 14.73 psia at 60°F. For prior years, the pressure base is 14.65 psia at 60°F.

Sources:

- 1949 through 1975, U.S. Department of the Interior, Bureau of Mines, Minerals Yearbook, "Natural Gas" chapter.
- 1976 through 1978, U.S. Department of Energy, Energy Information Administration, Natural Gas Production and Consumption, annual.

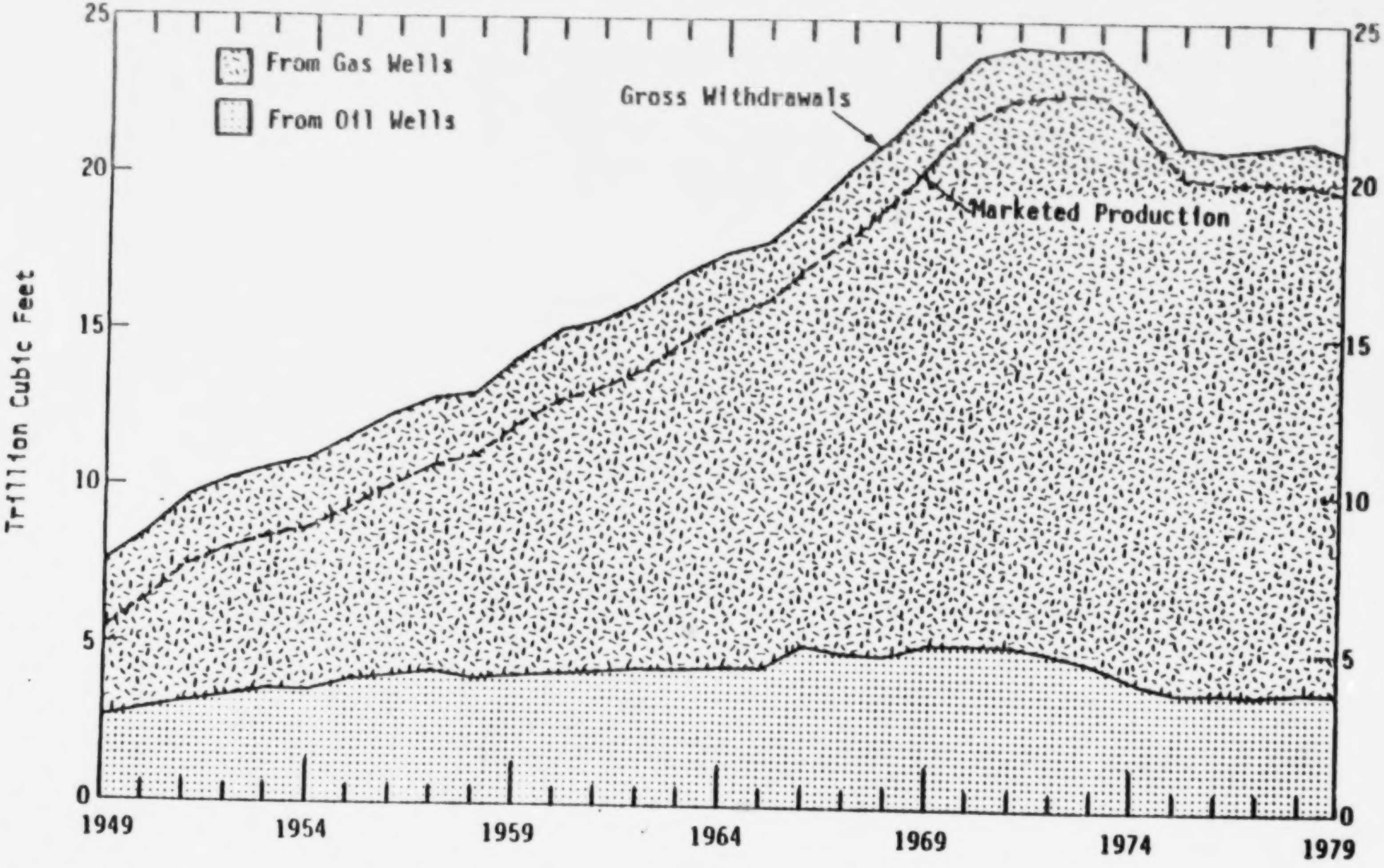

Figure 9-5. Natural gas gross withdrawals and marketed production.[12]

were nearly 157,000 producing gas wells in the United States. Although most natural gas is produced from natural gas wells, about 18 percent is produced from crude oil wells. Figure 9-6 portrays onshore and offshore natural gas production for the period 1954 through 1979.

The nominal price of natural gas remained reasonably steady for the long period from 1955 through 1970. Since 1973, the price has increased steadily in real terms. Figure 9-7 shows selected natural gas prices for three categories for the period 1955 through 1979. [2] In 1979, the price of natural gas at the wellhead was $1.13 per million Btu, $1.85 per million Btu at the city gate and $2.50 per million Btu delivered to the ultimate customer. Deregulation of the price of natural gas before the end of 1985 will boost the revenues and profitability margins for the industry. This will contribute to growth in capital availability which could potentially be used for more drilling, deeper drilling, and increased exploration and production of tight gas formations.

Since the Oil Embargo in 1973, the financial condition of the crude oil and natural gas production industry has been improving steadily in both revenues and net profits. Composite financial data shown in Table 9-14 reveals increased revenues, from $15,292 million in 1976 to $38,000 million in 1980. During the same period, net profits increased from $1,155 million to $1,925 million. Composite net profit margins as a percent of sales declined from 7.6 percent in 1976 to 5.1 percent in 1980. This fact indicates that production costs have risen at a faster pace than prices. Also, total capital has grown at a slower pace than revenues and profits. Consequently, return on total assets and return on equity have improved. According to Value Line Investment Survey, the composite industry will continue to have a healthy financial future into the 1980's. It is projected that in the period 1983 through 1985, the industry will have a composite net profit margin of 4.6 percent on annual revenues of approximately $70 billion, in current dollars. The long term debt ratio is projected to be 45.5 percent. Total capital is projected to increase to $35,500 million in current dollars or 51 percent of revenues in 1983-85. [13]

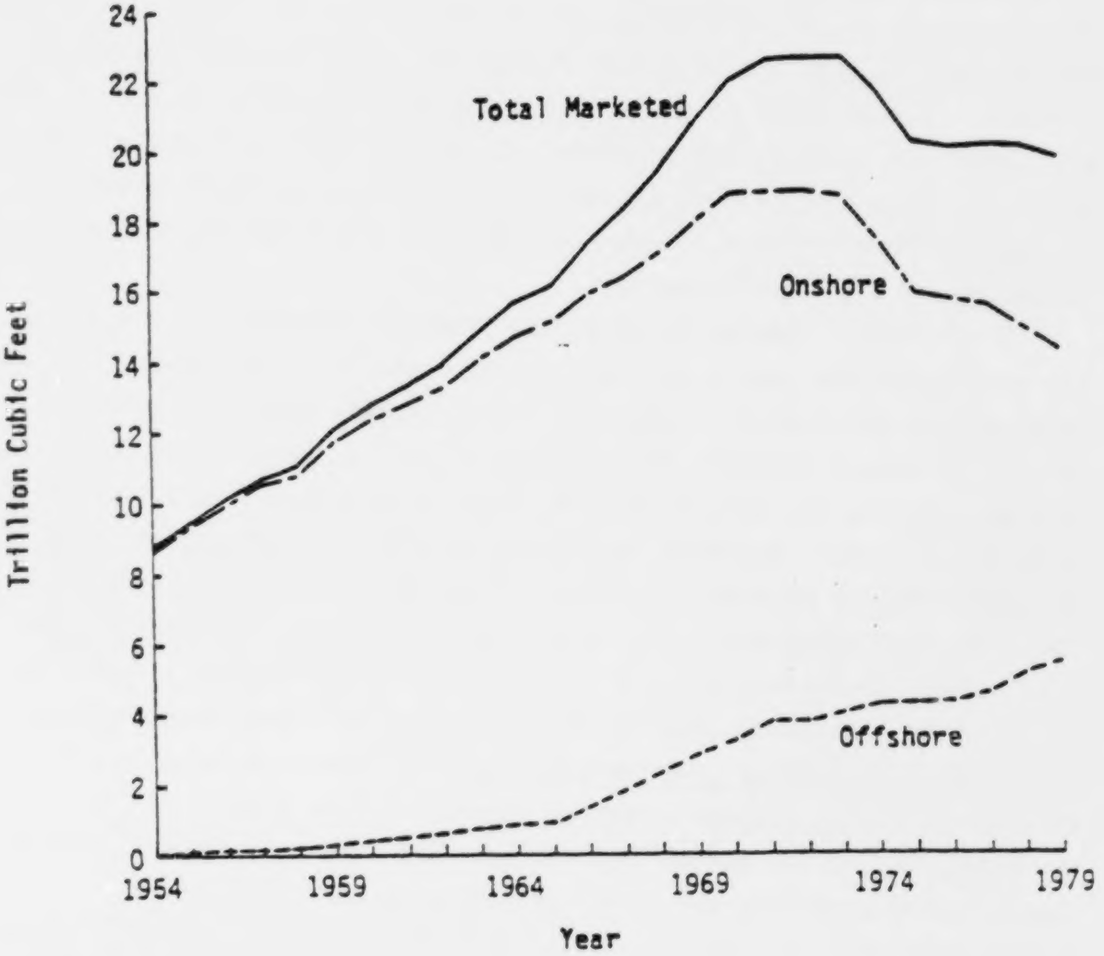

Figure 9-6. Onshore and offshore marketed natural gas production.[2]

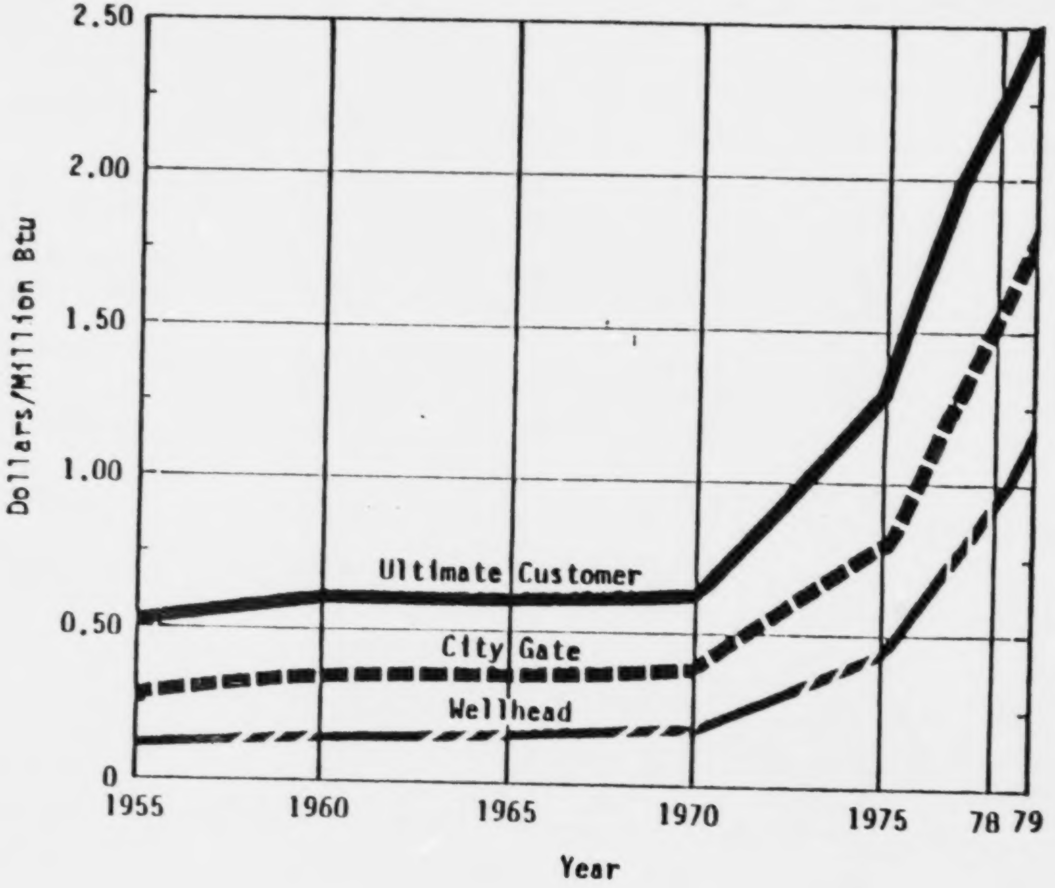

Figure 9-7. Selected natural gas prices – three categories for the period 1955-1979.[2]

Table 9-14. FINANCIAL DATA FOR THE NATURAL GAS INDUSTRY 1976-1981 and 1983-1985 ESTIMATES (Current dollars)

| Item | 1976 | 1977 | 1978 | 1979 | 1980 | 1981 | 83-85E |
|---|---|---|---|---|---|---|---|
| Revenues ($mill) | 15,292 | 19,430 | 22,463 | 30,357 | 38,000 | 46,000 | 70,000 |
| Net Profit ($mill) | 1,155 | 1,356 | 1,399 | 1,702 | 1,925 | 2,200 | 3,200 |
| Income Tax Rate | 44.4% | 43.1% | 43.9% | 43.2% | 44.0% | 45.0% | 47.0% |
| Net Profit Margin | 7.6% | 7.0% | 6.2% | 5.6% | 5.1% | 4.8% | 4.6% |
| Long-term Debt Ratio | 54.3 | 50.8 | 48.5 | 48.0 | 48.5 | 47.0 | 45.5 |
| Common Equity Ratio | 41.0 | 44.4 | 46.8 | 47.1 | 48.0 | 50.0 | 53.0 |
| Total Capital ($mill) | 19,538 | 20,207 | 20,611 | 22,236 | 23,750 | 26,000 | 35,500 |
| Net Plant ($mill) | 18,356 | 19,865 | 21,423 | 23,453 | 26,000 | 27,000 | 33,000 |
| % Earned Total Capital | 8.0% | 8.8% | 8.9% | 9.8% | 10.5% | 10.5% | 11.5% |
| % Earned Net Worth | 12.9% | 13.6% | 13.2% | 14.7% | 15.5% | 15.5% | 16.5% |
| % Earned Comm. Equity | 13.5% | 14.2% | 13.7% | 15.4% | 16.0% | 16.0% | 17.0% |
| % Retained to Comm. Equity | 7.5% | 8.0% | 7.2% | 8.9% | 9.0% | 9.0% | 9.5% |
| % All Dividends to Net Profit | 48% | 47% | 50% | 45% | 45% | 45% | 45% |
| Average Annual P/E Ratio | 7.1 | 7.6 | 7.1 | 6.8 | NA | NA | 8.0 |
| Average Annual Dividend Yield | 6.3% | 5.8% | 6.6% | 6.3% | NA | NA | 6.0% |
| Fixed Charge Coverage | 278% | 281% | 284% | 287% | 290% | 295% | 310% |

E = Estimates
NA = Not available

Source: A. Bernhard & Company. "Natural Gas Industry." Value Line Investment Survey, July 18, 1980.

Table 9-15 shows recovered sulfur production from natural gas production facilities and from petroleum refinery operations for the years 1960 through 1980.[9] For the period 1971 through 1980, recovered sulfur from natural gas plants accounted for 39 to 46 percent of the total recovered sulfur production. Sulfur recovered from petroleum refinery operations accounted for 54 to 61 percent. Separate data for natural gas plants and petroleum refineries for the years 1960 through 1970 were not available. Therefore, the production figures for these years are derived values based upon the 1971 through 1980 average of 43 percent of recovered sulfur from natural gas plants and 57 percent from petroleum refinery operations.

During the period from 1973 to 1979 when natural gas production remained relatively constant, the production of sulfur recovered from natural gas increased 68 percent from 1,063,000 megagrams in 1973 to 1,760,000 megagrams in 1979. A contributing factor to this growth was that sulfur recovery from natural gas was becoming economically attractive.

Sulfur prices during the last 27 years are shown in Table 9-16 in terms of actual and constant 1980 dollars per ton. Approximately 90 percent of all sulfur shipments are reflected in this table. Prices are based on the average reported rates for elemental sulfur (Frasch and recovered) f.o.b. mine or plant.

An ample supply of Frasch stocks resulted in a fairly stable sulfur market prior to 1964. Market prices rose in 1967 and 1968 due to the rapid growth in the fertilizer industry and a shortage of sulfur supply. In late 1968, a serious oversupply developed which led to a general collapse of the sulfur market through 1973.

There have been several factors contributing to dramatic sulfur price increases since 1974. Among these are: the rapid expansion of the fertilizer industry and its ability to pass on sulfur costs to farmers, logistic problems restricting delivery from other sources, higher production costs, and continued dependency upon Frasch sulfur.[10]

Table 9-17 presents a history of published prices for liquid sulfur at Tampa terminals in Florida for the period from 1969 through 1979. Tampa is one of several major Gulf ports where sulfur is traded in large volumes and market trends can be observed. Actual prices received by producers would depend on local sulfur markets or transportation costs to a suitable

Table 9-15. RECOVERED ELEMENTAL SULFUR PRODUCED IN THE UNITED STATES, 1960-1980

| Year | Total ($10^3$ Mg) | From natural gas plants ($10^3$ Mg) | Percentage of the total from natural gas plants | From petroleum refineries ($10^3$ Mg) | Percentage of the total from petroleum refineries |
|---|---|---|---|---|---|
| 1960 | 779 | 335[a] | 43.0 | 444[b] | 57.0 |
| 1961 | 872 | 375 | 43.0 | 497 | 57.0 |
| 1962 | 914 | 393 | 43.0 | 521 | 57.0 |
| 1963 | 962 | 413 | 43.0 | 549 | 57.0 |
| 1964 | 1,037 | 446 | 43.0 | 591 | 57.0 |
| 1965 | 1,234 | 530 | 43.0 | 704 | 57.0 |
| 1966 | 1,259 | 541 | 43.0 | 718 | 57.0 |
| 1967 | 1,288 | 554 | 43.0 | 734 | 57.0 |
| 1968 | 1,375 | 591 | 43.0 | 784 | 57.0 |
| 1969 | 1,437 | 618 | 43.0 | 819 | 57.0 |
| 1970 | 1,480 | 637 | 43.0 | 843 | 57.0 |
| 1971 | 1,620 | 648 | 40.0 | 972 | 60.0 |
| 1972 | 1,981 | 832 | 42.0 | 1,149 | 58.0 |
| 1973 | 2,455 | 1,063 | 43.3 | 1,392 | 56.7 |
| 1974 | 2,674 | 1,238 | 46.3 | 1,436 | 53.7 |
| 1975 | 3,017 | 1,364 | 45.2 | 1,653 | 54.8 |
| 1976 | 3,188 | 1,298 | 40.7 | 1,890 | 59.3 |
| 1977 | 3,624 | 1,426 | 39.3 | 2,198 | 60.7 |
| 1978 | 4,062 | 1,753 | 43.2 | 2,309 | 56.8 |
| 1979 | 4,070 | 1,760 | 43.2 | 2,310 | 56.8 |
| 1980 | 3,930 | 1,707 | 43.4 | 2,223 | 56.6 |

[a] Figures for the years 1960 through 1970 are derived values based on the assumption that sulfur recovered from natural gas is 43 percent of the total recovered sulfur production.

[b] Figures for the years 1960 through 1970 are derived values based on the assumption that sulfur recovered from petroleum refining operations is 57 percent of the total recovered sulfur production

Source: Bureau of Mines, Minerals Yearbook, 1960-1980 reprints.

Table 9-16. TIME-PRICE RELATIONSHIP FOR SULFUR, 1955-1981 [a]

| Year | Average annual price, dollars per metric ton | |
|---|---|---|
| | Actual prices | Based on constant 1980 dollars |
| 1955 | 27.50 | 80.21 |
| 1956 | 26.07 | 73.68 |
| 1957 | 24.02 | 65.65 |
| 1958 | 23.44 | 62.98 |
| 1959 | 23.09 | 60.61 |
| 1960 | 22.76 | 58.79 |
| 1961 | 22.75 | 58.23 |
| 1962 | 21.41 | 53.81 |
| 1963 | 19.67 | 48.70 |
| 1964 | 19.87 | 48.45 |
| 1965 | 22.12 | 52.79 |
| 1966 | 25.36 | 58.63 |
| 1967 | 32.12 | 72.09 |
| 1968 | 39.49 | 84.90 |
| 1969 | 26.62 | 54.43 |
| 1970 | 22.77 | 44.18 |
| 1971 | 17.19 | 31.77 |
| 1972 | 16.76 | 29.74 |
| 1973 | 17.56 | 29.48 |
| 1974 | 28.42 | 43.88 |
| 1975 | 44.91 | 63.47 |
| 1976 | 45.72 | 61.41 |
| 1977 | 44.38 | 56.32 |
| 1978 | 45.17 | 53.42 |
| 1979 | 55.75 | 60.78 |
| 1980 | 88.93 | 88.93 |
| 1981 | 111.00 P | NA |

NA - Not available

P Preliminary

[a] Frasch and recovered sulfur, f.o.b. mine/plant

Source: Bureau of Mines, Sulfur: Mineral Commodity Profile, 1979.

### Table 9-17. PUBLISHED PRICE FOR LIQUID SULFUR AT TAMPA TERMINALS

| Date | Price (current dollars) | |
|---|---|---|
| | Dollars per long ton | Dollars per metric ton |
| December 12, 1969 | 32.00 | 31.49 |
| February 1, 1970 | 30.00 | 29.53 |
| November 1, 1970 | 27.00 | 26.57 |
| March 10, 1971 | 25.00 | 24.60 |
| July 7, 1973 | 28.00 | 27.56 |
| September 12, 1973 | 31.00 | 30.51 |
| May 1, 1974 | 36.50 | 35.92 |
| June 6, 1974 | 42.00 | 41.34 |
| July 22, 1974 | 50.00 | 49.21 |
| January 1, 1975 | 57.00 | 56.10 |
| April 24, 1975 | 61.00 | 60.04 |
| December 1, 1975 | 65.00 | 63.97 |
| May 1, 1978 | 68.50 | 67.42 |
| February 9, 1979 | 73.25 | 72.09 |
| June 10, 1979 | 78.25 | 77.01 |
| September 1979 | 85.75 | 84.40 |
| November 1979 | 95.50 | 93.99 |
| March 1981 | 140.00 | 137.79 |

Source: Chemical Economies Handbook - Stanford Research Institute (SRI) International, December 1979 and Chemical and Engineering News Vol. 59 March 23, 1981, p. 25.

market, such as Tampa. The price has increased over 200 percent from $30.51 per metric ton in September 1973 to $93.99 per metric ton in November 1979. Historically, the sulfur price has been quite unstable. The March 1981 price was $140 per long ton f.o.b. Tampa.[6,14]

9.1.2.2 Five-Year Projections. In this subsection, projections for the number of new facilities constructed in the years 1983 through 1987 are developed. The size distribution of new facilities is developed based upon the industry's historical trend. Information on the projection of natural gas prices with deregulation are discussed.

Production of natural gas by conventional techniques has exceeded the rate of reserve additions in recent years. Consequently, conventional reserves and conventional production are expected to continue declining. Annual production of conventional natural gas is expected to decline roughly 1.5 to 2.0 trillion cubic feet every five years through 1995. The production of associated and dissolved gas in oil is expected to decline less rapidly than the production of nonassociated gas, due to higher price incentives for crude oil. Total domestic natural gas production is expected to continue declining.

Table 9-18 presents projected lower 48 states conventional natural gas production for the period 1980 through 2000.[15] In 1985, production is projected to be 19.7 trillion cubic feet, decreasing to 18.5 and 17.7 trillion cubic feet in 1987 and 1990, respectively.

Natural gas supply projections are conducted by various oil and gas companies as well as government and independent study groups. Table 9-19 presents a comparison of 1990 projection forecasts presented by the Department of Energy (DOE), the American Gas Association (AGA), Exxon, Tenneco, and other private study groups.[16] AGA's forecast of 16.3 quadrillion Btu per year is 2.8 percent lower than DOE's forecast of 17.8 quadrillion Btu per year, and Exxon's forecast of 14.9 quadrillion Btu is 16.3 percent lower than DOE's forecast. The AGA's forecast of projected lower 48 states conventional natural gas production was chosen for the present analyses for several reasons. AGA's forecast provides, as shown in Table 9-18, a detailed breakdown of production including newly discovered onshore and offshore categories. AGA's forecast also compares well with Exxon's, Tenneco's, and DOE's as indicated in Table 9-19. Projections for

## Table 9-18. PROJECTED LOWER-48 STATES CONVENTIONAL NATURAL GAS PRODUCTION, 1980-2000

| Gas source | Production, trillion standard cubic feet | | | | |
|---|---|---|---|---|---|
| | 1980 | 1985 | 1990 | 1995 | 2000 |
| Onshore | | | | | |
| Old inter[a] | 4.9 | 3.6 | 2.0 | 1.1 | 0.7 |
| Old intra[a] | 3.6 | 2.4 | 1.3 | 0.7 | 0.4 |
| Old direct sale | 4.0 | 2.6 | 1.5 | 0.8 | 0.5 |
| New | 1.5 | 3.6 | 4.9 | 4.8 | 3.8 |
| Offshore | | | | | |
| Old inter[a,b] | 5.6 | 4.1 | 1.4 | 0.7 | nil |
| New inter[c] | 0.1 | 3.4 | 6.6 | 6.5 | 5.4 |
| Total | | | | | |
| Old inter | 10.5 | 7.7 | 3.4 | 1.5 | 0.7 |
| Old intra | 3.6 | 2.4 | 1.3 | 0.7 | 0.4 |
| Old direct sale | 4.0 | 2.6 | 1.5 | 0.8 | 0.5 |
| New | 1.6 | 7.0 | 11.5 | 11.3 | 9.2 |
| TOTAL[d] | 19.7 | 19.7 | 17.7 | 14.6 | 10.8 |

[a] Includes gas used as compressor fuel and net storage injections.

[b] Including new additions from pre-1977 leases.

[c] Post-1976 leases only.

[d] Totals may not add due to independent rounding.

Source: American Gas Association, Gas Supply and Statistics, "Total Energy Resource Analysis Model (TERA) 80-1," Appendix A. Figure A-2, p.21.

Table 9-19. PROJECTIONS OF NATURAL GAS SUPPLY: COMPARISON OF 1990 FORECASTS (Quadrillion Btu)

| | | 1979 Projections for 1990 | | | | | |
|---|---|---|---|---|---|---|---|
| Units | 1978 Actual | DOE/EIA[a] | AGA[b] | DPP[c] | Pace[d] | Exxon[e] | Tenneco[f] |
| Domestic Production | | | | | | | |
| Conventional | 19.5 | 17.8 | 15.3-17.3 | 16.9 | 16.1 | 14.9 | 14.8 |
| North Alaska | g | 0.9 | 1.6 | 0.4 | 1.0 | g | 1.0 |
| Synthetic Gas | 0.2 | 0.3 | 1.1 | 0.6 | 0.8 | 0.6-1.0 | 1.5 |
| Subtotal | 19.7 | 19.0 | 19.9-21.9 | 18.0 | 18.0 | 15.5-15.9 | 17.3 |
| Net Imports | | | | | | | |
| Pipeline | 0.9 | 0 | 2.1 | 2.0 | 1.4 | 1.8 | 2.0 |
| Liquefied Natural Gas | h | 0.8 | 2.0 | 1.0 | 0.8 | 0.8 | 3.1 |
| Subtotal | 0.9 | 0.8 | 4.2 | 3.1 | 2.2 | 2.7 | 5.1 |
| Total Supply | 20.6 | 19.8 | 24.1-26.1 | 21.0 | 20.2 | 18.2-18.6 | 22.4 |

[a] DOE/EIA 1979 Annual Report to Congress, middle range forecast.

[b] American Gas Association, The Future for Gas Energy in the United States, June 1979.

[c] Data Resources, Inc., Energy Review, Winter 1980.

[d] The Pace Company Consultants and Engineers, Inc., The Pace Energy and Petrochemical Outlook to 2000, October 1979.

[e] Exxon Company, U.S.A., Energy Outlook 1980-2101, December 1979.

[f] Tenneco Oil Company, Energy 1979-2000, June 1979.

[g] Included in conventional production.

[h] Less than 0.5 quadrillion Btu.

Note: Non-EIA projections converted from trillion cubic feet with 1,020 Btu per cubic foot. Numbers may not add to totals because of rounding.

the amount of sulfur that will be recovered from natural gas were not available. However, it is expected that with the increased new onshore discoveries of natural gas and increased deep well drilling activities, recovery of elemental sulfur will increase and will continue to be economically attractive. Full sulfur recovery capacity utilization is expected within the range of limiting factors, such as the sour gas supply rate.

The projection of the number of new onshore natural gas sweetening and sulfur recovery facilities was based upon:

(1) historical data on the sulfur intake capacity (sulfur feed rate) distribution of the existing 89 onshore natural gas sulfur recovery facilities,

(2) the projected new discovery of onshore natural gas for the period 1983-1987,

(3) the projected sulfur composition of the newly discovered gas, and

(4) the distribution among various plant sizes of the projected sulfur production from newly discovered gas and, therefore, the number and size distribution of new facilities.

The following factors were also considered in making these projections:

(1) The combined sulfur intake capacity of the 89 existing sulfur recovery facilities is approximately 10,640 megagrams per day (Table 9-5).

(2) Historically, the percent capacity utilization of these facilities has ranged from as high as 88 percent to as low as 47 percent (Table 9-8).

(3) Recovered sulfur produced in the U.S. has consistently increased from 335,000 megagrams in 1960 to 1,707,000 megagrams in 1980 (Table 9-15).

(4) The price of liquid sulfur has increased from $32 per megagram in December 1969 to $94 per megagram in November 1979 (Table 9-17).

9.1.2.2.1 Sulfur Intake Capacity Distribution. Table 9-9 indicates that the percent distribution of sulfur intake capacity of all existing facilities at the end of 1972 is the same as the percent distribution at the end of 1982. The Oil and Gas Journal lists recent worldwide construction of 22 sulfur recovery units with a size distribution of 63 percent at 0-34 Mg/d, 23 percent at 35-74 Mg/d and 14 percent at greater than 500 Mg/d. The Oil and Gas Journal's distribution compares well with the 1972 and 1982 distributions among the existing 89 facilities. Therefore, the 1982 distribution was considered the basis for the distribution of the projected sulfur in newly discovered gas among various plant sizes.

9.1.2.2.2 Projected New Onshore Natural Gas Production. Projected new onshore natural gas production for each individual year for 1983-1987, was derived from the American Gas Association's (AGA) data (Table 9-18). This derivation is presented herewith in a tabular form (Table 9-20). An average of 6.2 percent per year depletion was assumed for new discoveries. 26
The projected new production discovered in each year is as indicated in Table 9-21.

9.1.2.2.3 Projected Sulfur in Newly Discovered Onshore Natural Gas. Sulfur (as $H_2S$) in sour natural gas, that is further processed for sulfur recovery, has increased at a consistent rate. Table 9-22 indicates that the sulfur content increased from 1.32 mole percent (average) in 1960 to 5.92 mole percent (average) in 1980. The sulfur content values in Table 9-21 were derived based on the following: (1) 25 percent of total onshore production in 1982 was sour and, therefore, was sweetened and (2) three-fourths of the sour gas was further processed for sulfur recovery.[17,18] This represents an average of 18.75 percent of the total onshore production that was processed for sulfur recovery in the 1982. These sulfur content values in sour natural gas were obtained from the 1982 API survey. To estimate sulfur in new gas, it is assumed that 25 percent of the total onshore production will be sour and three-fourths of the sour gas will be further processed for sulfur recovery.[22,23,24] That is, an average of 18.75 percent of the total new onshore discovery will be processed for sulfur recovery. Therefore, the remainder (6.25 percent) will be incinerated without sulfur recovery. This view is supported by two facts: (1) increased deep well drilling activities will be enhanced by price deregulation; and (2) the increasing price of recovered sulfur compounded with the decreased supply from Frasch mines. Also, with the increased price of recovered sulfur, it becomes more economically feasible to recover sulfur. However, in the absence of any reliable information or

## Table 9-20. DERIVATION OF NEWLY DISCOVERED ONSHORE CONVENTIONAL NATURAL GAS PRODUCTION FOR A SPECIFIED YEAR, 1977-1987

| | Cumulative Production (tscf) | New onshore production, trillion standard cubic feet (tscf) | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Year | | 1977 | 1978 | 1979 | 1980 | 1981 | 1982 | 1983 | 1984 | 1985 | 1986 | 1987 |
| 1977 | 0.27 | 0.27 | | | | | | | | | | |
| 1978 | 0.68 | 0.25 | 0.43 | | | | | | | | | |
| 1979 | 1.09 | 0.238 | 0.403 | 0.449 | | | | | | | | |
| 1980 | 1.50 | 0.223 | 0.378 | 0.421 | 0.478 | | | | | | | |
| 1981 | 1.92 | 0.209 | 0.355 | 0.395 | 0.448 | 0.513 | | | | | | |
| 1982 | 2.34 | 0.196 | 0.333 | 0.371 | 0.421 | 0.481 | 0.538 | | | | | |
| 1983 | 2.76 | 0.184 | 0.312 | 0.348 | 0.395 | 0.451 | 0.505 | 0.565 | | | | |
| 1984 | 3.18 | 0.172 | 0.293 | 0.326 | 0.370 | 0.423 | 0.473 | 0.530 | 0.593 | | | |
| 1985 | 3.60 | 0.162 | 0.275 | 0.306 | 0.347 | 0.397 | 0.444 | 0.497 | 0.556 | 0.616 | | |
| 1986 | 3.86 | 0.152 | 0.258 | 0.287 | 0.326 | 0.372 | 0.416 | 0.466 | 0.522 | 0.578 | 0.483 | |
| 1987 | 4.12 | 0.142 | 0.242 | 0.269 | 0.305 | 0.349 | 0.391 | 0.437 | 0.489 | 0.542 | 0.453 | 0.501 |

Note: (1) "Newly discovered" means that produced after April 20, 1977.

(2) Column 2 indicates cumulative new discovered production from April 20, 1977 onwards. For example, the 1980's cumulative new production of 1.5 tscf is the production beyond April 20, 1977 through 1980.

(3) Newly discovered production for an individual year is indicated by the last number in the row for a specified year. For example, 1980's individual newly discovered production (produced in 1980 only) is 0.478 tscf. A new discovery depletion rate of 6 percent is applied to derive individual newly discovered production for a specified year. For example, the newly discovered production in 1978 only is equal to 1978's cumulative production of 0.68 tscf minus 94 percent of 1977's individual production of 0.27 tscf, that is 0.68 tscf minus 0.25 tscf. This gives 0.43 tscf as individual new production for 1978.

Table 9-21. PROJECTED NEW SULFUR RECOVERY CAPACITIES FOR THE PERIOD OF 1983-1987

| Year | New Onshore natural gas production in the specified year MMcf | Onshore production that is sour[a] MMcf | Throughput for sweetening only[b] MMcf | Throughput for sulfur recovery[c] MMcf | Average $H_2S$ in the sour natural gas streams processed for sulfur recovery, mole percent | New sulfur recovery capacity added in the specified year $10^3$ Mg/y | New sulfur recovery capacity added in the specified year[d], Mg/d |
|---|---|---|---|---|---|---|---|
| 1983 | 565,000 | 141,250 | 36,380 | 104,870 | 5.8 | 232.2 | 663.4 |
| 1984 | 593,000 | 148,250 | 38,180 | 110,070 | 5.8 | 244.1 | 697.4 |
| 1985 | 616,000 | 154,000 | 39,670 | 114,330 | 5.8 | 253.3 | 723.7 |
| 1986 | 483,000 | 120,750 | 31,100 | 89,650 | 5.8 | 199.2 | 569.1 |
| 1987 | 501,000 | 125,250 | 32,260 | 92,990 | 5.8 | 206.5 | 590.0 |
| Total in period of 1983-1987 | 2,758,000 | 689,500 | 177,590 | 511,910 | | 1,135 | 3,244 |

[a] Derived values based upon an average of 25 percent of the onshore production in the period of 1983-1987 being sour and the remaining 75 percent being sweet.

[b] Derived values based upon 25.7 percent of the sour onshore new discovery for the projected period 1983-1987 processed for sweetening only.

[c] Derived values based upon 74.3 percent of the sour onshore new discovery for the projected period 1983-1987 processed for sulfur recovery.

[d] 350 days/year operating schedule.

Table 9-22. SULFUR RECOVERED PER UNIT VOLUME OF PROCESSED SOUR NATURAL GAS, 1954-1980

| Year | Onshore natural gas production, tscf | Onshore production processed for sulfur recovery[a], tscf | Sulfur recovery from natural gas, $10^3$ Mg | Sulfur recovered[b], kg/Mcf sour natural gas | Sulfur (as $H_2S$)[b] in sour natural gas processed for sulfur recovery, mole percent |
|---|---|---|---|---|---|
| 1954 | 8.66 | .548 | N.A.[c] | –[d] | – |
| 1955 | 9.28 | .587 | N.A. | – | – |
| 1956 | 9.94 | .629 | N.A. | – | – |
| 1957 | 10.51 | .665 | N.A. | – | – |
| 1958 | 10.77 | .682 | N.A. | – | – |
| 1959 | 11.70 | .741 | N.A. | – | – |
| 1960 | 12.33 | .780 | 335 | 0.431 | 1.32 |
| 1961 | 12.77 | .808 | 375 | 0.463 | 1.43 |
| 1962 | 13.24 | .838 | 393 | 0.467 | 1.44 |
| 1963 | 13.99 | .886 | 413 | 0.467 | 1.43 |
| 1964 | 14.70 | .931 | 446 | 0.481 | 1.47 |
| 1965 | 15.10 | .956 | 530 | 0.553 | 1.71 |
| 1966 | 15.84 | 1.003 | 541 | 0.539 | 1.66 |
| 1967 | 16.33 | 1.034 | 554 | 0.535 | 1.65 |
| 1968 | 17.00 | 1.076 | 591 | 0.549 | 1.69 |
| 1969 | 17.86 | 1.131 | 618 | 0.549 | 1.68 |
| 1970 | 18.70 | 1.184 | 637 | 0.539 | 1.66 |
| 1971 | 18.74 | 1.186 | 648 | 0.544 | 1.68 |
| 1972 | 18.77 | 1.188 | 832 | 0.699 | 2.15 |

(continued)

Table 9-22. Continued

| Year | Onshore natural gas production, tscf | Onshore production processed for sulfur recovery[a], tscf | Sulfur recovery from natural gas, $10^3$ Mg | Sulfur recovered[b], kg/Mcf sour natural gas | Sulfur (as $H_2S$)[b] in sour natural gas processed for sulfur recovery, mole percent |
|---|---|---|---|---|---|
| 1973 | 18.67 | 1.182 | 1,063 | 0.898 | 2.77 |
| 1974 | 17.37 | 1.100 | 1,238 | 1.125 | 3.46 |
| 1975 | 15.85 | 1.003 | 1,364 | 1.361 | 4.18 |
| 1976 | 15.65 | .991 | 1,298 | 1.311 | 4.03 |
| 1977 | 15.49 | .991 | 1,426 | 1.456 | 4.47 |
| 1978 | 14.87 | .941 | 1,753 | 1.864 | 5.73 |
| 1979 | 14.25 | .902 | 1,760 | 1.950 | 6.00 |
| 1980 | 14.00 | .886 | 1,707 | 1.928 | 5.92 |

[a] Based upon 19 percent of the total onshore production for the period 1960-1980 being sour and the remaining 81 percent being sweet. Thirty-three percent of the sour gas for the period 1960-1980 is further processed for sulfur recovery.

[b] Derived values based upon average sulfur recovery efficiency of 85 percent.

[c] Data on recovered sulfur from the onshore natural gas produced prior to 1960 is not available.

[d] Not applicable.

data, the sulfur content (as $H_2S$) in new onshore discoveries was assumed to remain the same as the current weighted average of 5.8 mole percent (Appendix G, Table G-3). The sulfur content value (weighted average) for the period 1983 through 1987 and the projected new sulfur recovery capacity is indicated in Table 9-21.

9.1.2.2.4 Distribution of New Sulfur in Various Plant Sizes. The projected sulfur in new onshore discoveries for the period 1983 through 1987 was distributed among various plant sizes according to the 1982 plant size distribution. It was assumed that all new discoveries will require construction and operation of new sulfur recovery facilities. This resulted in 30 facilities processing gas for sulfur recovery during the period 1983 through 1987. The sulfur (0.2 mole percent as shown in Appendix G, Table G-2) in the portion of the new gas that is sweetened and incinerated without sulfur recovery was divided among the remaining facilities with capacities that range from less than 0.1 megagrams per day to 5.1 megagrams per day. These small facilities are described in Appendix G, Tables G-5 and G-7. This resulted in 37 facilities that sweeten sour gas but do not recover sulfur. Projected new sulfur recovery facilities are presented in Table 9-23.

The projected volume of new onshore natural gas in 1980 and 1987, by category relevant to this analysis, is summarized in Table 9-24. In 1987, domestic production is projected to be 18.5 trillion cubic feet. Including imports and total domestic natural gas, total natural gas supply is expected to be approximately 20 trillion cubic feet. New onshore natural gas production is projected to be 2.8 trillion cubic feet. Twenty-five percent of this is assumed to be sour, which is equivalent to .70 trillion cubic feet of gas requiring sweetening at new facilities. The total volume of new onshore natural gas associated with sulfur recovery in new facilities is projected to be three-fourths of .70 trillion cubic feet or .53 trillion cubic feet. This will represent approximately 2.7 percent of the total estimated domestic natural gas supply of 20 trillion cubic feet in 1987. Thus, in 1987, the proposed $SO_2$ NSPS could apply to only 2.7 percent of the total domestic natural gas supply. This figure places an

Table 9-23. PROJECTED ADDED NEW SULFUR RECOVERY FACILITIES AND THEIR SULFUR INTAKE SIZE DISTRIBUTION, 1983-1987

| Year | (Avg. size 10), Mg/d | (Avg. size 102), Mg/d | (Avg. size 564), Mg/d | (Avg. size 1,016), Mg/d | Total |
|---|---|---|---|---|---|
| 1983 | 3 | 1 | – | – | 4 |
| 1984 | 4 | 2 | 1 | – | 7 |
| 1985 | 4 | 3 | – | 1 | 8 |
| 1986 | 4 | 2 | – | – | 6 |
| 1987 | 3 | 1 | 1 | – | 5 |
| Number | 18 | 9 | 2 | 1 | 30 |
| Percent of total number | 60 | 30 | 6.7 | 3.3 | 100 |
| New sulfur recovery capacity added, Mg/d | 180 | 918 | 1,128 | 1,016 | 3,244 |

## Table 9-24. ESTIMATED NATURAL GAS PRODUCTION, ONSHORE VOLUME, SWEETENED VOLUMES, VOLUMES ASSOCIATED WITH SULFUR RECOVERY, 1980 AND 1987

| Category | 1980 | Baseline 1987 (est.) |
|---|---|---|
| | (tcf) | (tcf) |
| Natural gas production (lower 48 states) | 19.7 | 18.5 |
| Onshore natural gas production | 14.0 | 11.1 [a] |
| Onshore natural gas treated for $H_2S$ removal | 2.7 [b] | 2.8 [c] |
| Onshore natural gas associated with sulfur recovery | .90 | .93 |
| New onshore natural gas production | e | 2.8 |
| New onshore natural gas treated for $H_2S$ removal in new facilities [b] | e | .70 [c] |
| New onshore natural gas associated with sulfur recovery in new facilities [c] | e | .53 [d] |

---

[a] 1987 estimates assumed 60 percent of domestic supply is onshore.

[b] Nineteen percent of onshore gas is sour in 1980.

[c] Twenty-five percent of onshore gas will be sour.

[d] Seventy-five percent of new sour natural gas is projected to be treated for sulfur recovery.

[e] Not applicable.

upper bound on any supply quantity impact of these regulations in 1987. Furthermore, considering the price elasticity of domestic retail natural gas demand (-.426), the maximum potential retail price impact is less than a 5 percent price increase. These impacts are further clarified in the following sections of this chapter.

By 1985, almost all categories of natural gas production will be deregulated. Very little new gas will be subject to controls; most old intrastate gas will be decontrolled, and the quantity of old interstate gas that remains controlled will decline rapidly over time. Natural gas prices are projected by the U.S. Department of Energy to increase due of the Natural Gas Policy Act and phased deregulation of prices during the period 1983 through 1987. Deregulated prices are expected to boost exploration and production activities. The history and projections of natural gas prices are summarized in Table 9-25. [21] In 1987, the average wellhead price of new natural gas, including intrastate and interstate, is interpolated to be $4.80 per Mcf in 1980 dollars.

## Table 9-25. NATURAL GAS PRICES: HISTORY AND PROJECTIONS FOR 1965-1995
(1980 Dollars per Thousand Cubic Feet)[d]

| Price | History[a] | | | Projection[b] | | |
|---|---|---|---|---|---|---|
| | 1965 | 1973 | 1978 | 1985 | 1990 | 1995 |
| Domestic Wellhead Prices | | | | | | |
| Old Interstate | NA | NA | 1.01 | 1.10 | 1.29 | 1.52 |
| New Interstate | NA | NA | NA | 4.88 | 4.40 | 5.00 |
| Old Intrastate | NA | NA | NA | 3.59 | 3.62 | 4.12 |
| New Intrastate | NA | NA | NA | 5.15 | 4.67 | 5.25 |
| North Alaska | -- | -- | -- | -- | 2.02 | 2.02 |
| Average | 0.39 | 0.38 | 1.11 | 3.55 | 3.73 | 4.55 |
| Synthetic Gas Prices | | | | | | |
| High-Btu Coal Gas | -- | -- | -- | 5.19 | 4.57 | 5.13 |
| Medium-Btu Coal Gas | -- | -- | -- | 4.03 | 4.91 | 5.93 |
| Imported Gas Prices | | | | | | |
| Canadian Gas | NA | NA | 2.63 | 6.77 | 7.54 | 9.28 |
| Mexican Gas | NA | NA | NA | 6.77 | 7.54 | 9.28 |
| Liquefied Natural Gas | -- | -- | 1.68 | 6.44 | 7.00 | 8.39 |
| Delivered Prices | | | | | | |
| Residential | 2.55 | 2.22 | 3.02 | 5.90 | 6.26 | 7.03 |
| Commercial | 1.74 | 1.59 | 2.59 | 5.32 | 5.69 | 6.46 |
| Raw Material | NA | NA | NA | 4.67 | 4.88 | 5.68 |
| Large boilers[c] | NA | NA | NA | 5.71 | 4.95 | 5.73 |
| Industrial, Other | 0.85 | 0.84 | 1.76 | 4.73 | 4.92 | 5.69 |
| Refineries | NA | NA | NA | 4.96 | 4.83 | 5.59 |
| Electric Utilities | 0.97 | 0.69 | 1.88 | 5.17 | 4.82 | -- |
| Alternative Fuel Cost | -- | -- | -- | 6.79 | 7.57 | 9.04 |

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[a] Source for historical data is Volume 2 of the EIA Annual Report to Congress, 1979, and the following EIA Energy Data Reports: Natural Gas Production and Consumption, 1978; United States Imports and Exports of Natural Gas, 1978; and, Natural and Synthetic Gas, 1978.

[b] Projections for the middle oil price scenario.

[c] Major fuel-burning installations.

[d] Inflated by GNP implicit price deflator, 9.0 percent from 1979 dollars to 1980 dollars.

Notes: NA = Not available.
-- = Not applicable.

Source: DOE/EIA Annual Report to Congress, 1980, Vol. 13, pg. 90.

## 9.2 ECONOMIC IMPACT ANALYSIS

This section contains a discussion of the economic analysis methodology used to assess the impacts of alternative regulations controlling sulfur dioxide ($SO_2$) emissions from sour natural gas sweetening and sulfur recovery processes. In addition, estimates of these economic impacts are presented.

Total additional before-tax annualized costs of controls in 1987, the fifth year of controls, for the projected sixty-seven new natural gas sulfur recovery plants range from zero under Regulatory Alternatives I and II to $98.7 million under Regulatory Alternative VI. Aggregate after-tax annualized emissions control costs amount to less than one-half of one percent of the total projected value of new natural gas production in 1987 under Regulatory Alternatives I through VI. Thus, the impacts of $SO_2$ emissions regulations on expected returns from natural gas exploration and development are relatively small and the effects on exploration for and development of new natural gas fields are expected to be negligible.

Unit emissions control costs serve as a useful indicator of profitability impacts per Mcf on individual regulated plants, assuming no natural gas price impacts. As shown in Table 9-26, across all model plants, unit emissions control costs range from zero under Regulatory Alternative I to $10.48 per Mcf under Regulatory Alternative VI. In general, and assuming that the effects of the regulations on the exploration for and development of new natural gas are negligible, natural gas production and price impacts are expected to be negligible under Regulatory Alternatives I through VI. Under Regulatory Alternatives IV, V and VI, additional emissions control costs are estimated to cause less than 9 small natural gas sulfur recovery plants to be unable to fully recover total production costs (see Table 9-27). It is further estimated that one to two of these nine plants will likely be curtailed under Regulatory Alternatives V-VI because unit variable costs (including emissions control costs) will exceed the forecasted price of natural gas. The decline in natural gas production associated with these potential curtailments would be negligible.

| Model plant | Size (sulfur intake) Mg/d | Size (sulfur intake) LT/d | Acid gas feed $H_2S/CO_2$ ratio by volume | Regulatory alternative II | III | IV | V | VI |
|---|---|---|---|---|---|---|---|---|
| 1 | <0.1 | <0.1 | 12.5/87.5 | 0 | 0 | 0 | 0 | 0 |
| 2 | 0.1 | 0.1 | 12.5/87.5 | 0 | 0 | 0 | 0 | 0 |
| 3 | 0.2 | 0.2 | 12.5/87.5 | 0 | 0 | 0 | 0 | .66 to 10.48 |
| 4 | 0.3 | 0.3 | 12.5/87.5 | 0 | 0 | 0 | .44 to 6.99 | .44 to 6.99 |
| 5 | 0.4 | 0.4 | 12.5/87.5 | 0 | 0 | 0 | .33 to 5.24 | .39 to 6.20 |
| 6 | 0.5 | 0.5 | 12.5/87.5 | 0 | 0 | .26 to 4.19 | .31 to 4.96 | .38 to 6.09 |
| 7 | 0.6 | 0.6 | 12.5/87.5 | 0 | 0 | .22 to 3.49 | .26 to 4.14 | .32 to 5.08 |
| 8 | 0.7 | 0.7 | 12.5/87.5 | 0 | 0 | .19 to 2.99 | .22 to 3.54 | .27 to 4.35 |
| 9 | 0.8 | 0.8 | 12.5/87.5 | 0 | 0 | .16 to 2.62 | .19 to 3.10 | .24 to 3.81 |
| 10 | 0.9 | 0.9 | 12.5/87.5 | 0 | 0 | .15 to 2.33 | .17 to 2.76 | .21 to 3.38 |
| 11 | 1 | 1 | 12.5/87.5 | 0 | .13 to 2.10 | .13 to 2.10 | .16 to 2.48 | .19 to 3.05 |
| 12 | 2 | 2 | 12.5/87.5 | 0 | .07 to 1.05 | .07 to 1.05 | .08 to 1.24 | .10 to 1.52 |
| 13 | 3 | 3 | 12.5/87.5 | 0 | .04 to .70 | .05 to .83 | .05 to .83 | .06 to 1.02 |
| 14 | 4 | 4 | 12.5/87.5 | 0 | .03 to .52 | .04 to .62 | .04 to .62 | .05 to .76 |
| 15 | 5.1 | 5 | 12.5/87.5 | 0 | .03 to .42 | .03 to .50 | .03 to .50 | .04 to .61 |
| 16 | 10.2 | 10 | 50/50 | 0 | .00 to .01 | .00 to .01 | .01 to .13 | .01 to .13 |
| 17 | 10.2 | 10 | 12.5/87.5 | 0 | .00 to .02 | .00 to .02 | .01 to .20 | .01 to .20 |
| 18 | 101.6 | 100 | 50/50 | 0 | .01 to .03 | .01 to .03 | .14 to .53 | .14 to .53 |
| 19 | 101.6 | 100 | 20/80 | 0 | .01 to .05 | .01 to .05 | .21 to .85 | .21 to .85 |
| 20 | 563.8 | 555 | 50/50 | [a] | [a] | [a] | [a] | [a] |
| 21 | 563.8 | 555 | 80/20 | 0 | .02 to .09 | .02 to .09 | .02 to .09 | .02 to .09 |
| 22 | 1,015.9 | 1,000 | 50/50 | [a] | [a] | [a] | [a] | [a] |
| 23 | 1,015.9 | 1,000 | 80/20 | 0 | .02 to .07 | .02 to .07 | .02 to .07 | .02 to .07 |

Regulatory Alternative I is the baseline.

[a] Not estimated. No plants are projected for this model plant.

## Table 9-27. INCREASES IN THE CUMULATIVE NUMBER OF ONSHORE SOUR NATURAL GAS SULFUR RECOVERY PLANTS UNABLE TO FULLY COVER TOTAL EXPLORATION AND PRODUCTION COSTS DUE TO REGULATORY ALTERNATIVES, 1987

| Model plant | Regulatory alternative | | | |
|---|---|---|---|---|
| | II-III | IV | V | VI |
| 1 | 0 | 0 | 0 | 0 |
| 2 | 0 | 0 | 0 | 0 |
| 3 | 0 | 0 | 0 | 1 |
| 4 | 0 | 0 | <1 | <1 |
| 5 | 0 | 0 | <1 | <1 |
| 6 | 0 | <1 | <1 | <1 |
| 7 | 0 | <1 | <1 | <1 |
| 8 | 0 | <1 | <1 | <1 |
| 9 | 0 | <1 | <1 | <1 |
| 10 | 0 | <1 | <1 | <1 |
| 11 | 0 | <1 | <1 | <1 |
| 12 | 0 | 0 | 0 | 0 |
| 13 | 0 | 0 | 0 | 0 |
| 14 | 0 | 0 | 0 | 0 |
| 15 | 0 | 0 | 0 | 0 |
| 16 | 0 | 0 | 0 | 0 |
| 17 | 0 | 0 | 0 | 0 |
| 18 | 0 | 0 | 0 | 0 |
| 19 | 0 | 0 | 0 | 0 |
| 20 | 0 | 0 | 0 | 0 |
| 21 | 0 | 0 | 0 | 0 |
| 22 | 0 | 0 | 0 | 0 |
| 23 | 0 | 0 | 0 | 0 |

Regulatory Alternative I is the baseline.
< - "less than"

Sulfur supplies are expected to increase slightly as a result of regulations which require increased sulfur recovery efficiency. Under Regulatory Alternatives IV, V and VI, natural gas sulfur recovery would increase by approximately 0.1 million metric tons. This represents about a 0.5 percent increase in total projected sulfur supply in 1987 and could cause a slight decrease in local market sulfur prices.

The remainder of Chapter 9 is organized as follows. A discussion of the impact assessment methodology is presented in Section 9.2.1. Section 9.2.2 presents a detailed discussion of the estimated economic impacts of each $SO_2$ regulatory alternative. Section 9.3 describes the potential socioeconomic and inflationary impacts.

### 9.2.1 Economic Impact Assessment Methodology

Each regulatory alternative will have two effects on the decisions of firms to produce onshore natural gas from new fields. First, by increasing the expected total costs of exploring for and processing onshore natural gas, the regulatory alternatives reduce the expected returns to firms from exploration for and exploitation of new natural gas fields 1/. Thus, firms will be likely to reduce levels of exploration activities, thereby reducing the number of fields that will be discovered. Second, by increasing the costs of obtaining gas from discovered sour natural gas fields, regulatory alternatives may reduce the production of gas from those fields. In this analysis, the impacts of the regulatory alternatives on the rate of exploration and discovery of new fields are assessed qualitatively. The effects of the regulatory alternatives on the rate of production of natural gas from discovered fields are assessed quantitatively.

The methodology used in the analysis is as follows. Initially, it is assumed that the baseline forecasts of the number and size of potential natural gas sulfur recovery plants presented in Table 9-23 are unaffected by each regulatory alternative. This amounts to the assumption that the

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1/ To the extent that exploration activities are non-directional (that is, concerned with the discovery of hydrocarbons in the form of oil and/or natural gas) the effect of the regulation on exploration activities will be diminished.

effects of each standard on expected returns to exploration for and exploitation of new natural gas fields are negligible and therefore that these activities will not be curtailed as a result of the regulatory alternatives. The validity of this assumption for each regulatory alternative is examined carefully. The methodology used to carry out this assessment is described later in this section.

Second, the percent $H_2S$ in sour natural gas is estimated for each model plant projected to be constructed under the baseline scenario. These projections of the quality of natural gas in each plant are obtained using the data presented in Table 9-7 and pertain to sour gas that is subject to sweetening and sulfur recovery. These data account for the forecasted 5.8 percent $H_2S$ in sour natural gas between 1980 and 1987 (presented in Table 9-21) 1/. It is necessary to estimate the distribution of plants by $H_2S$ percentage and sulfur capacity in order to calculate the level of sweet gas output for each model plant. This estimate is required to calculate unit emissions control costs.

Third, long run baseline average total costs of producing sweetened natural gas from new wells in 1987 (measured in 1980 constant dollars) are estimated. The estimation procedure is as follows. Data on new well onshore natural gas production costs for 1980 were obtained from the API and DOE. These data are presented in Table 9-4. These cost estimates are adjusted to reflect costs of production in 1987 under the assumption that the average total costs of new gas production will, in the absence of regulation, rise to $4.80 per Mcf from the 1980 level of $2.00 per Mcf. This forecasted rise in real production costs is predicated on the assumption that the natural gas industry is competitive and that in the long run, average total costs of production from new wells will be equal to the expected price of natural gas. In this case, the expected price is assumed to be the forecasted deregulated natural gas price of $4.80 per

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1/ Between 1980 and 1987, the average proportion of $H_2S$ by volume in sour natural gas is expected to remain about 5.8 percent. The distribution of plants by percent $H_2S$ natural gas presented in Table 9-7 is qualitatively adjusted to take account of this projected $H_2S$ concentration to obtain the distribution of plants by $H_2S$ concentration presented in Tables 9-33 through 9-55.

Mcf 1/. Total costs of production vary from well to well as a result of differences in well depths and gas flow rates. These differences may be substantial, ranging from $.05 per Mcf to $72.18 per Mcf in terms of 1980 dollars. Clearly, if a firm knew before the fact that the unit total costs of producing natural gas from an individual well would exceed average revenues from the sale of that gas (that is, the market price of natural gas), it would not drill the well. However, in general, until the well has been drilled and is in operation, the firm does not know what its unit total costs of production will be. Consequently, once the well has been drilled, the firm bases its decision on whether or not it will operate the well, on whether or not unit variable costs of production exceed average revenues (i.e., the well-head price). Unit variable costs of production of natural gas are relatively small when compared to unit total costs, ranging from a minimum of $.03 per Mcf to a maximum of $.93 per Mcf in terms of 1980 constant dollars. (See Table 9-4, Column 4.) Note that unit variable costs of production include field equipment costs and operating and maintenance costs. These are costs that can be avoided by firms if they decide not to produce gas from a given field, once the depth of the field and well sizes have been determined. Further note that unit variable costs, defined as costs that could be avoided once the nature of the well has been determined, also include all costs associated with compliance with each regulatory alternative because these costs could also be avoided by the firm if it decided not to produce gas from a particular well. The methodology used to compute before-tax and after-tax costs of pollution control for each model plant is presented in Section 9.2.1.1. In the case of non-processing gas producing firms that sell sour gas to processors, the cost of compliance would be reflected in increased charges by the processing company to cover the costs associated with the regulatory alternatives, or alternatively, an equivalent reduction in the well head price paid for the sour gas. In the case of gas producing/processing companies, costs of compliance would be directly incurred.

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1/ This forecast price was obtained from DOE and is measured in terms of constant 1980 dollars.

The above discussion provides a simple decision rule for the firm with respect to whether or not it will choose to produce and process gas from any given well. If price exceeds unit variable cost then the well will be utilized to produce natural gas. Under the assumed baseline, all of the 67 facilities forecasted to be constructed will be used to process natural gas as the forecasted deregulated price of $4.80 per Mcf exceeds the maximum forecasted unit variable cost of $0.93 per Mcf. Curtailments will occur as a result of each regulatory alternative if and only if unit pollution control costs exceed the difference between the predicted price and pre-regulation unit variable costs, given the assumption that the impacts of the regulatory alternative on exploration and exploitation of new fields are negligible.

The extent to which each regulatory alternative is likely to affect exploration and drilling activities is assessed in two ways. First, the impact of each regulatory alternative on the expected number of facilities that will recover, or more than recover, all costs of production is calculated. These data provide a qualitative measure of the extent to which the regulatory alternatives will reduce expected returns from exploration for and exploitation of potential natural gas fields. Second, the total costs of each regulatory alternative relative to the value of total new gas production are computed. The methodology used to estimate the effect of each regulatory alternative on potential profitability of the exploration and plant processing operations, predicted under the baseline scenario, requires further discussion. This discussion is presented in 9.2.1.2.

The above analysis is carried out under the assumption that the well head price for sweet gas will be unaffected by any of the regulatory alternatives. The basis for this assumption is as follows. Sour natural gas cannot be sold at sweet natural gas prices. Either it is priced below sweet gas or is not marketed. The main reason for a differential between sour and sweet natural gas prices, according to industry sources, is that $H_2S$ in natural gas is corrosive to compressors, valves and other pipeline and natural gas processing equipment. Pipeline companies are therefore reluctant to accept natural gases with a high $H_2S$ concentration.

Consequently, sour gas must be sweetened, i.e., the $H_2S$ must be removed, or blended with sweet natural gases in order to be marketable as sweet gas according to industry standards.

Sweetening is performed by either producers, gas processors or pipeline companies. These operations may be vertically integrated or independent operations depending on local market conditions. It is common practice for gas processors and pipeline companies to charge producers and dealers (third parties) a treating fee for sulfur removal against sour gas. This practice pushes sweetening costs backward to the producer. The producer then chooses either to sell sour gas at the sour gas price, to install and operate his own sweetener and sell sweetened natural gas or to cap his sour gas producing well(s). The alternative of capping the sour gas well(s) may be a temporary measure until a sufficient number of exploration wells are drilled and enough gas discovered to justify further field development costs.

Under existing and expected market conditions, producers and their leaseholders (royalty owners) bear the cost of sweetening their produced sour natural gases. Regulations which require additional emissions control costs on new sweetening and new sulfur recovery operations in the onshore natural gas production industry will effectively increase sour gas sweetening costs in new plants and producers and leaseholders of new sour gas will thus incur lower net incomes. Individually, sour gas producers are not expected to be able to pass-on additional emissions control costs because this would require sweetened natural gas to sell at a premium over naturally sweet gas. Historically, producers have been unable to obtain such a premium and none is expected in the future. Thus, if natural gas prices were to increase due to $SO_2$ emissions regulations, they would only increase as a result of plant curtailments and associated supply shifts. But, as shown in Table 9-24, even in the unlikely event that all new sour natural gas sulfur recovery plants and their associated natural gas production were curtailed, the loss of domestic natural gas supply would not exceed .51 trillion cubic feet, less than 3 percent of total domestic natural gas consumption. The expected number of curtailments under the most stringent regulatory alternative is much less severe, less than one tenth of one percent of projected domestic consumption, implying a negligible increase in domestic prices.

9.2.1.1 **Annualized and Unit Emissions Control Costs**. Before-tax annualized costs (BTAC) and after-tax annualized costs (ATAC) of emissions controls are computed for each model plant and regulatory alternative using the following equations:

$$BTAC = I_0 \; CRF + O\&M_0 \qquad (1)$$

$$ATAC = I_0 \; CRF \; TAXF + (1-t) \; O\&M_0 \qquad (2)$$

where,

$I_0$ = initial base year investment

$O\&M_0$ = annual operating & maintenance costs less applicable by-product (sulfur and steam) credits

$CRF = \frac{r(1+r)^n}{(1+r)^n-1}$, the capital recovery factor

$r$ = the real cost of capital = $1 - \frac{1+d}{1+inf}$

$n$ = economic life of the asset, i.e. the capital recovery period (variable by asset)

$TAXF = 1 - itc - t \; PVDEP$

$itc$ = investment tax credit rate

$t$ = corporate income tax rate

$PVDEP$ = present value of annual depreciation factors per $1 of investment, i.e.

$$PVDEP = \sum_{y=1}^{Y} \frac{DEP_y}{(1+d)^y}$$

$Y$ = length of the depreciation period, 3, 5, 10 or 15 years

$d$ = nominal discount rate (the weighted nominal cost of capital)

$DEP_y$ = annual depreciation factors calculated using the most advantageous depreciation methods for the firm, either (1) rapid amortization of pollution control investments or (2) accelerated cost recovery as allowed by the 1981 Economy Recovery Act.

Inflation and the weighted, nominal, after-tax cost of capital are projected to be 8 and 10 percent, respectively. The inflation rate is based on recent estimates obtained from the DRI econometric model of the U.S. economy.[25] The nominal weighted after-tax cost of capital for the natural gas industry was based on a composite natural gas industry stock price earnings ratio of 7 to 8. In addition, the marginal corporate income tax rate was assumed to be 47 percent, the debt ratio was assumed to be 45 percent, and the nominal pre-tax interest rate on new debt for domestic corporations was estimated to be 13 percent.[13]

The real industry cost of capital (value of r) was computed to be .019. This is based on the estimated nominal discount rate (d=.10) and the inflation rate (inf = .08); $r = 1-(1+d)/(1+inf)$. Thus the real industry cost of capital, based on 1980 industry and financial market conditions, is .019. This is a different rate than used in Chapter 8. The rate used in Chapter 8 is a constant 10 percent used by the government to make interindustry cost-effectiveness comparisons. The rate used in Chapter 9 depends on financial and market conditions and is appropriate for capital budgeting and economic impact analysis.

Unit emissions control costs for each model plant, measured in 1980 dollars per Mcf of sweet gas, are computed using the following equation.

$$\text{Unit emissions control cost} = \frac{\text{ATAC}}{\text{Sweet gas volume}} \qquad (3)$$

where ATAC is the after-tax annualized cost of emission controls as defined in Equation 2. Sweet gas volume is a function of the sulfur intake of the natural gas sulfur recovery plant and the $H_2S$ and $H_2S/CO_2$ ratios in the sour natural gas (See Table 9-28, Footnote a).

The estimated 1987 baseline sweet gas sales and sulfur sales associated with each model plant and regulatory alternative are shown in Tables 9-28 and 9-29, respectively. Natural gas and sulfur volumes associated with each level of sales can be derived by simply dividing each level of sales by the appropriate baseline price per unit. For natural gas, the baseline price is $4.80 per Mcf. For sulfur, the baseline price is $98.43 per Mg.

9.2.1.2 Long-run Profitability of Regulated Plants. In the context of this analysis, a new onshore natural gas sulfur recovery plant and its associated natural gas production is defined to be profitable in the long run if the wellhead gas price is greater than or equal to its actual unit production costs for drilling and equipping wells (including dry holes). These costs include field equipment, operating and maintenance expenditures, sweetening costs and baseline net sulfur recovery revenues. In post regulation scenarios, incremental $SO_2$ emission control costs are also included in addition to exploration and development drilling costs.

Note that this definition of new plant long run profitability takes into account the recovery of total production costs, including a return on investment. Therefore, in the long run a plant is unprofitable if it fails to recover total production cost as defined above. It is important to recognize, however, that although a particular well may fail to yield revenues that cover all production costs, it will never-the-less be operated if all variable costs of production are covered (that is, as long as the price of natural gas exceeds average variable costs). However, a well will not be operated if average variable operating costs exceed the price of natural gas. Under this scenario, a plant can make no contribution to the recovery of exploration and drilling costs.

Under the baseline scenario, 54 of the 67 projected sulfur recovery plants will be profitable in the long run, and thirteen will not. All 67 sulfur plants, however, will be operated because their variable unit costs are smaller than the predicted wellhead price for sweet gas of $4.80 per Mcf. The estimation procedure used to calculate the number of natural gas sulfur recovery operations that, under the baseline scenario, would cover all production costs is as follows. A probability distribution of total

Table 9-28. MODEL PLANT ESTIMATED SWEET GAS SALES[a]

| Model plant | $H_2S$ fraction in sour natural gas | | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | 0.005 | 0.010 | 0.020 | 0.030 | 0.040 | 0.050 | 0.060 | 0.070 | 0.080 | 0.090 | 0.100 | 0.150 | 0.200 |
| | million 1980 dollars | | | | | | | | | | | | |
| 1 | 0.3 | 0.1 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | NA | NA |
| 2 | 0.9 | 0.4 | 0.2 | 0.1 | 0.1 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 | NA | NA |
| 3 | 1.7 | 0.8 | 0.4 | 0.2 | 0.2 | 0.1 | 0.1 | 0.1 | 0.0 | 0.0 | 0.0 | NA | NA |
| 4 | 2.6 | 1.2 | 0.6 | 0.3 | 0.2 | 0.2 | 0.1 | 0.1 | 0.1 | 0.0 | 0.0 | NA | NA |
| 5 | 3.4 | 1.6 | 0.7 | 0.5 | 0.3 | 0.2 | 0.2 | 0.1 | 0.1 | 0.1 | 0.0 | NA | NA |
| 6 | 4.3 | 2.1 | 0.9 | 0.6 | 0.4 | 0.3 | 0.2 | 0.1 | 0.1 | 0.1 | 0.0 | NA | NA |
| 7 | 5.1 | 2.5 | 1.1 | 0.7 | 0.5 | 0.3 | 0.2 | 0.2 | 0.1 | 0.1 | 0.1 | NA | NA |
| 8 | 6.0 | 2.9 | 1.3 | 0.8 | 0.5 | 0.4 | 0.3 | 0.2 | 0.1 | 0.1 | 0.1 | NA | NA |
| 9 | 6.9 | 3.3 | 1.5 | 0.9 | 0.6 | 0.4 | 0.3 | 0.2 | 0.2 | 0.1 | 0.1 | NA | NA |
| 10 | 7.7 | 3.7 | 1.7 | 1.0 | 0.7 | 0.5 | 0.3 | 0.3 | 0.2 | 0.1 | 0.1 | NA | NA |
| 11 | 8.6 | 4.1 | 1.9 | 1.1 | 0.8 | 0.5 | 0.4 | 0.3 | 0.2 | 0.1 | 0.1 | NA | NA |
| 12 | 17.1 | 8.2 | 3.7 | 2.3 | 1.5 | 1.1 | 0.8 | 0.6 | 0.4 | 0.2 | 0.2 | NA | NA |
| 13 | 25.7 | 12.3 | 5.6 | 3.4 | 2.3 | 1.6 | 1.2 | 0.8 | 0.6 | 0.4 | 0.3 | NA | NA |
| 14 | 34.3 | 16.4 | 7.5 | 4.5 | 3.0 | 2.1 | 1.5 | 1.1 | 0.8 | 0.6 | 0.4 | NA | NA |
| 15 | 42.8 | 20.5 | 9.4 | 5.7 | 3.8 | 2.7 | 1.9 | 1.4 | 1.0 | 0.7 | 0.4 | NA | NA |
| 16 | 88.4 | 43.7 | 21.4 | 14.0 | 10.3 | 8.0 | 6.5 | 5.5 | 4.7 | 4.1 | 3.6 | NA | NA |
| 17 | 85.7 | 41.1 | 18.7 | 11.3 | 7.6 | 5.4 | 3.9 | 2.8 | 2.0 | 1.4 | 0.9 | NA | NA |
| 18 | 883.7 | 437.4 | 214.2 | 139.8 | 102.7 | 80.3 | 65.5 | 54.8 | 46.9 | 40.7 | 35.7 | 20.8 | 13.4 |
| 19 | 870.3 | 424.0 | 200.8 | 126.5 | 89.3 | 66.9 | 52.1 | 41.4 | 33.5 | 27.3 | 22.3 | 7.4 | 0.0 |
| 20 | 4904.5 | 2427.5 | 1189.0 | 776.1 | 569.7 | 445.9 | 363.3 | 304.3 | 260.1 | 225.7 | 198.2 | 115.6 | 74.3 |
| 21 | 4923.1 | 2446.1 | 1207.6 | 794.7 | 588.3 | 464.4 | 381.9 | 322.9 | 278.7 | 244.3 | 216.7 | 134.2 | 92.9 |
| 22 | 8837.0 | 4373.9 | 2142.3 | 1398.4 | 1026.5 | 803.4 | 654.6 | 548.3 | 468.6 | 406.6 | 357.1 | 208.3 | 133.9 |
| 23 | 8870.5 | 4407.3 | 2175.8 | 1431.9 | 1060.0 | 836.8 | 688.1 | 581.8 | 502.1 | 440.1 | 390.5 | 241.8 | 167.4 |

NA = Not applicable

[a] Sweet gas sales equals sweet gas volume times price. Price is assumed to be \$4.70/MMBtu (\$4.80/Mcf) (the Department of Energy's 1987 projected well head price in 1980 dollars). Sweet gas volume (bcfy) equals $.0092982 \frac{S}{c}\left[1 - c - \frac{c}{r}\right]$ where S = sulfur intake (LT/D), c = $H_2S$ fraction in sour natural gas, and r = $H_2S/CO_2$ ratio.

## Table 9-29. MODEL PLANT ESTIMATED 1987 VALUE OF SULFUR SALES [a], BY REGULATORY ALTERNATIVES

| Model plant | Size (sulfur intake) Mg/d | Size (sulfur intake) LT/d | Acid gas feed $H_2S/CO_2$ ratio by volume | Regulatory alternative I Baseline control level | II | III | IV | V | VI |
|---|---|---|---|---|---|---|---|---|---|
| | | | | Million 1980 dollars | | | | | |
| 1-15 | $\leq$5.1 | $\leq$5 | b | .00[c] | .00[d] | .00[d] | .00[d] | .00[d] | .00[d] |
| 16 & 17 | 10.2 | 10 | b | .328 | .328 | .326 | .336 | .341 | .341 |
| 18 & 19 | 101.6 | 100 | b | 3.361 | 3.361 | 3.444 | 3.444 | 3.496 | 3.496 |
| 20 | 563.8 | 555 | 50/50 | 18.654 | 19.114 | 19.114 | 19.412 | 19.412 | 19.412 |
| 21 | 563.8 | 555 | 80/20 | 18.771 | 19.132 | 19.420 | 19.420 | 19.420 | 19.420 |
| 22 | 1,015.9 | 1,000 | 50/50 | 33.615 | 34.445 | 34.445 | 34.980 | 34.980 | 34.980 |
| 23 | 1,015.9 | 1,000 | 80/20 | 33.822 | 34.473 | 34.991 | 34.991 | 34.991 | 34.991 |

[a] Sulfur price is \$98.43/Mg (\$100/LT). Operation is 350 days per year.

[b] Covers the entire ratio from 12.5/87.5 to 80/20.

[c] Sulfur is not recovered in the baseline.

[d] Sulfur sales are less than \$175,000 per year and are assumed to be zero for purposes of estimating after-tax net annualized emissions control costs.

unit production costs of natural gas from new sour wells was estimated utilizing the distribution of gas production by well size (Table 9-1) and the distribution of the numbers of new wells drilled by depth (Table 9-2). It is assumed that existing well sizes (flow rates) reflect expected new well sizes and that well size and well depth are independent. The marginal and bivariate distributions of well depth, flow rate and the combination of well depth and flow rate thus obtained are shown in Table 9-30.

The probability estimates are created in order to approximate the distribution of 1979-1980 natural gas production costs to the wellhead from new wells. The distribution indicates that, in 1979-80 total unit production costs associated with production from new wells, averaged over the entire distribution, were approximately $2.00 per Mcf.

For the reasons discussed in Section 9.2.1 above, by 1987, average total unit production costs are assumed to increase to $4.80 in terms of 1980 constant dollars. It is also assumed that total unit costs of production rise in proportion to the increase in average total unit costs across the entire distribution of production costs of gas from new wells.

Utilizing these assumptions, the following probability distribution for the average total cost of producing natural gas from new sour gas wells was obtained.

| Cost of Production (1980$/Mcf sweet gas) | Estimated probability |
|---|---|
| less than 2.40 | .60 |
| 2.40 to 3.60 | .10 |
| 3.60 to 4.80 | .10 |
| 4.80 and above | .20 |

Approximately 20 percent of all natural gas produced from new sour gas wells will have total unit operating costs in excess of $4.80 per Mcf. Sixty percent will have total unit production costs below $2.40 per Mcf. Approximately 10 percent will have total costs between $2.40 and $3.60 per Mcf and an additional 10 percent will have total costs between $3.60 and $4.80 per Mcf. Given the above distribution for average total costs of production, under the baseline scenario, 80% or 54 of the 67 projected natural gas sulfur recovery facilities and their associated natural gas production processes will cover all production costs while 13 will not.

Table 9-30. ESTIMATED ONSHORE NATURAL GAS PRODUCTION COSTS AND PROBABILITY ESTIMATES FOR SELECTED WELL DEPTHS AND BASE YEAR FLOW RATES, 1980

| | | | | Well depth (feet) | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | | 2,000 | | 4,000 | | 8,000 | | 12,000 | |
| Drilling and equipment cost per productive gas well [a] (1980$): | | | | 89,000 | | 211,000 | | 949,000 | | 1,875,000 | |
| P(d) [d] | | | | .30 | | .50 | | .10 | | .10 | |
| Flow rate (Mcfd) | P(r) | Field equipment cost [b, c] (1980$) | Annual O&M cost [b] (1980 $/yr) | Unit total cost and (P(r,d)) [d] 1980 $/Mcf | | | | | | | |
| 50 | .02 | 15,000 | 6,000 ($.36/Mcf) | 4.31 | (.006) | 8.95 | (.01) | 36.99 | (.002) | 72.18 | (.002) |
| 250 | .20 | 26,000 | 12,000 ($.14/Mcf) | 1.02 | (.060) | 1.95 | (.10) | 7.55 | (.020) | 14.59 | (.020) |
| 500 | .36 | 35,000 | 15,000 ($.09/Mcf) | .56 | (.108) | 1.02 | (.18) | 3.83 | (.036) | 7.35 | (.036) |
| 1,000 | .40 | 43,000 | 20,000 ($.06/Mcf) | .31 | (.120) | .54 | (.20) | 1.94 | (.040) | 3.70 | (.040) |
| 5,000 | .02 | 64,000 | 29,000 ($.02/Mcf) | .08 | (.006) | .12 | (.01) | .40 | (.002) | .75 | (.002) |
| 10,000 | .00 | 87,000 | 44,000 ($.01/Mcf) | .05 | (.000) | .07 | (.00) | .21 | (.000) | .39 | (.000) |

[a] Source is Table 9-3.

[b] Source: "Aggregate Average Annual Gas Well Operating Costs vs Production Capacity 1981" DOE, Dallas, TX, (1981 figures deflated to 1980 by 9 percent) (See Reference 5.)

[c] Includes field equipment between "Christmas tree" and point of transfer. Does not include sour gas sweetening or sulfur recovery equipment.

[d] P(x) - probability estimates for well depth (d), flow rate (r), and flow rate and well depth (r,d).

The projected distribution of 1987 wellhead production costs from new sour gas wells is used to estimate the corresponding distribution of long run baseline unit producers' surplus. Unit producer surplus is measured as price less baseline long run unit production cost per Mcf. The resulting preregulation distribution of unit producer's surplus is summarized below:

| Wellhead price | Production costs | Baseline unit producers' surplus | Estimated probability |
|---|---|---|---|
| ------------(1980 $/Mcf sweet gas)------------ | | | |
| 4.80 | less than 2.40 | 2.40 to 4.80 (3.60) | .60 |
| 4.80 | 2.40 to 3.60 | 1.20 to 2.40 (1.80) | .10 |
| 4.80 | 3.60 to 4.80 | 0 to 1.20 ( .60) | .10 |
| 4.80 | 4.80 and above | less than 0.0 | .20 |

Note: The numbers in parenthesis indicate the midpoint of the range.

The distribution of baseline producer surplus presented above indicates the level of additional emissions control costs that new sulfur recovery plants can incur and continue to cover all production costs. Thus, if all plants were located at the mid-point of the range of the above probability distribution for producer surplus, then 10 percent of the new onshore natural gas production from sour natural gas would become unprofitable in the long run if incremental $SO_2$ emissions control costs exceeded $.60 per Mcf. Approximately 20 percent would become unprofitable in the long run if incremental control costs exceeded $1.80 per Mcf. Approximately 80 percent remain profitable even if incremental control costs were as high as $2.40 per Mcf.

In order to estimate the numbers of each type of model plant that are expected to cease to cover total production costs as a result of each regulatory alternative, the proportions of each type of model plants expected to have negative producer surpluses were estimated. To obtain measures of the actual number of plants that are likely to become unprofitable in the long run, these proportions were multiplied by the baseline projections of the numbers of each type of model plants. In order to assess the effect of each regulatory alternative on long run industry profitability, the number of plants expected to be unprofitable in the long run under each regulatory alternative is compared with the number of plants projected to be unprofitable in the long run under the baseline scenario.

### 9.2.2 Economic Impacts of $SO_2$ NSPS Regulatory Alternatives on Sour Gas Sweetening and Sulfur Recovery Plants

This section presents the economic impacts that are associated with each regulatory alternative.

9.2.2.1 Net Annualized Costs Per Plant. The incremental before-tax and after-tax net annualized costs for individual model plants associated with Regulatory Alternatives I through VI are shown in Tables 9-31 and 9-32, respectively. These costs were computed using the methodology presented in Section 9.2.1.1 and provide a basis for further economic impact analysis. A comparison of before-tax and after-tax net annualized costs provides a measure of the income tax consequences of the regulatory alternatives to the firm, assuming that the firm has taxable net income and does not pass additional emissions control costs forward to consumers or backward to suppliers. The results indicate that almost 50 percent of the before-tax costs of emissions control will be borne indirectly by the federal government through investment tax credits and depreciation and expense deductions.

9.2.2.2 Unit Emission Control Costs, Curtailments and Long Run Profitability of Model Plants. The relative impact of each regulatory alternative on sour gas sweetening and sulfur recovery plants is indicated by the magnitude of the emissions control cost per Mcf of associated sweet gas output. Assuming no price impacts, the unit emissions control costs also indicate profitability impacts per Mcf. The unit emissions control costs for twenty-three model plants, six regulatory alternatives and thirteen different $H_2S$ percentages in sour natural gas are shown in Tables 9-33 through 9-55. The tables also present the estimated distribution of model plants according to the predicted $H_2S$ content of sour natural gas.

The unit emissions control costs are compared to the 1987 wellhead price of $4.80 per Mcf for new natural gas, to the projected onshore natural gas production costs from new sour gas wells, and to the projected baseline unit producer surplus. If unit emissions control costs exceed $.60, they are marked with one asterisk (*) indicating a 10 percent estimated probability that any given model plant would cease to cover total production costs due to the regulatory alternative. If unit emission

Table 9-31. MODEL PLANT BEFORE-TAX ANNUALIZED COST, BY REGULATORY ALTERNATIVE

| Model plant | Size (sulfur intake) |  | Acid gas feed $H_2S/CO_2$ ratio by volume | Regulatory alternative |  |  |  |  |
|---|---|---|---|---|---|---|---|---|
|  | Mg/d | Lt/d |  | II | III | IV | V | VI |
|  |  |  |  | Thousands of 1980 dollars |  |  |  |  |
| 1 | <0.1 | <0.1 | a | 0 | 0 | 0 | 0 | 0 |
| 2 | 0.1 | 0.1 | a | 0 | 0 | 0 | 0 | 0 |
| 3 | 0.2 | 0.2 | a | 0 | 0 | 0 | 0 | 428 |
| 4 | 0.3 | 0.3 | a | 0 | 0 | 0 | 428 | 428 |
| 5 | 0.4 | 0.4 | a | 0 | 0 | 0 | 428 | 506 |
| 6 | 0.5 | 0.5 | a | 0 | 0 | 428 | 506 | 600 |
| 7 | 0.6 | 0.6 | a | 0 | 0 | 428 | 506 | 600 |
| 8 | 0.7 | 0.7 | a | 0 | 0 | 428 | 506 | 600 |
| 9 | 0.8 | 0.8 | a | 0 | 0 | 428 | 506 | 600 |
| 10 | 0.9 | 0.9 | a | 0 | 0 | 428 | 506 | 600 |
| 11 | 1 | 1 | a | 0 | 428 | 428 | 506 | 600 |
| 12 | 2 | 2 | a | 0 | 428 | 428 | 506 | 600 |
| 13 | 3 | 3 | a | 0 | 428 | 506 | 506 | 600 |
| 14 | 4 | 4 | a | 0 | 428 | 506 | 506 | 600 |
| 15 | 5.1 | 5 | a | 0 | 428 | 506 | 506 | 600 |
| 16-17 | 10.2 | 10 | a | 0 | 44 | 44 | 388 | 388 |
| 18-19 | 101.6 | 100 | a | 0 | 452 | 452 | 7,352 | 7,352 |
| 20 | 563.8 | 555 | 50/50 | 142 | 142 | 9,581 | 9,581 | 9,581 |
| 21 | 563.8 | 555 | 80/20 | -56 | 3,374 | 3,374 | 3,374 | 3,374 |
| 22 | 1,015.9 | 1,000 | 50/50 | -119 | -119 | 15,316 | 15,316 | 15,316 |
| 23 | 1,015.9 | 1,000 | 80/20 | -7 | 4,718 | 4,718 | 4,718 | 4,718 |

Regulatory Alternative I is the baseline.

[a] Covers the entire ratio range from 12.5/87.5 to 80/20.

## Table 9-32. MODEL PLANT AFTER-TAX ANNUALIZED COST, BY REGULATORY ALTERNATIVE

| Model plant | Size (sulfur intake) |  | Acid gas feed $H_2S/CO_2$ ratio by volume | Regulatory alternative |  |  |  |  |
|---|---|---|---|---|---|---|---|---|
|  | Mg/d | LT/d |  | II | III | IV | V | VI |
|  |  |  |  | ------------- | ------------- | Thousands of 1980 dollars | ------------- | ------------- |
| 1 | <0.1 | <0.1 | a | 0 | 0 | 0 | 0 | 0 |
| 2 | 0.1 | 0.1 | a | 0 | 0 | 0 | 0 | 0 |
| 3 | 0.2 | 0.2 | a | 0 | 0 | 0 | 0 | 234 |
| 4 | 0.3 | 0.3 | a | 0 | 0 | 0 | 234 | 234 |
| 5 | 0.4 | 0.4 | a | 0 | 0 | 0 | 234 | 277 |
| 6 | 0.5 | 0.5 | a | 0 | 0 | 234 | 277 | 340 |
| 7 | 0.6 | 0.6 | a | 0 | 0 | 234 | 277 | 340 |
| 8 | 0.7 | 0.7 | a | 0 | 0 | 234 | 277 | 340 |
| 9 | 0.8 | 0.8 | a | 0 | 0 | 234 | 277 | 340 |
| 10 | 0.9 | 0.9 | a | 0 | 0 | 234 | 277 | 340 |
| 11 | 1 | 1 | a | 0 | 234 | 234 | 277 | 340 |
| 12 | 2 | 2 | a | 0 | 234 | 234 | 277 | 340 |
| 13 | 3 | 3 | a | 0 | 234 | 277 | 277 | 340 |
| 14 | 4 | 4 | a | 0 | 234 | 277 | 277 | 340 |
| 15 | 5.1 | 5 | a | 0 | 234 | 277 | 277 | 340 |
| 16-17 | 10.2 | 10 | a | 0 | 24 | 24 | 221 | 221 |
| 18-19 | 101.6 | 100 | a | 0 | 245 | 245 | 3,940 | 3,940 |
| 20 | 563.8 | 555 | 50/50 | 84 | 84 | 5,125 | 5,125 | 5,125 |
| 21 | 563.8 | 555 | 80/20 | -25 | 1,834 | 1,834 | 1,834 | 1,834 |
| 22 | 1,015.9 | 1,000 | 50/50 | -52 | -52 | 8,185 | 8,185 | 8,185 |
| 23 | 1,015.9 | 1,000 | 80/20 | 1 | 2,555 | 2,555 | 2,555 | 2,555 |

Regulatory Alternative I is the baseline.

[a] Covers the entire ratio range from 12.5/87.5 to 80/20.

TABLE 9-33. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 1 ( 0.0345 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE |  |  |  |  |
|---|---|---|---|---|---|---|
|  |  | 2 | 3 | 4 | 5 | 6 |
|  |  | 1980 $/MCF |  |  |  |  |
| 0.5 | 3.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| 1.0 | 2.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| 2.0 | 2.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| 3.0 | 1.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| 4.0 | 1.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| 5.0 | 1.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| 6.0 | 1.0 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
|  | 11. |  |  |  |  |  |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

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TABLE 9-34. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 2 ( 0.1016 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE | | | | |
|---|---|---|---|---|---|---|
| | | 2 | 3 | 4 | 5 | 6 |
| | | 1980 $/MCF | | | | |
| 0.5 | 0.3 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| 1.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| 2.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| 3.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| 4.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| 5.0 | 0.1 | 0.0 | 0.0 | 0.0 | 0.0 | 0.0 |
| | 1. | | | | | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

TABLE 9-35. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 3 ( 0.2032 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| | | 1980 $/MCF | | | | |
| 0.5 | 0.9 | 0.0 | 0.0 | 0.0 | 0.0 | 0.50 * |
| 1.0 | 0.6 | 0.0 | 0.0 | 0.0 | 0.0 | 1.37 * |
| 2.0 | 0.6 | 0.0 | 0.0 | 0.0 | 0.0 | 3.30 *** |
| 3.0 | 0.3 | 0.0 | 0.0 | 0.0 | 0.0 | 4.97 **** |
| 4.0 | 0.3 | 0.0 | 0.0 | 0.0 | 0.0 | 7.40 **** |
| 5.0 | 0.3 | 0.0 | 0.0 | 0.0 | 0.0 | 10.48 **** |
| | 3. | | | 0.0 | 0.0 | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

* THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 10 PERCENT.

** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 20 PERCENT.

*** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS GREATER THAN 20 PERCENT.

**** THE PLANT WOULD BE CURTAILED.

TABLE 9-36. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 4 ( 0.3048 MC/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE | | | | |
|---|---|---|---|---|---|---|
| | | 2 | 3 | 4 | 5 | 6 |
| | | 1980 $/MCF | | | | |
| 0.5 | 0.3 | 0.0 | 0.0 | 0.0 | 0.44 | 0.44 |
| 1.0 | 0.2 | 0.0 | 0.0 | 0.0 | 0.91 * | 0.91 * |
| 2.0 | 0.2 | 0.0 | 0.0 | 0.0 | 2.00 ** | 2.00 ** |
| 3.0 | 0.1 | 0.0 | 0.0 | 0.0 | 3.31 *** | 3.31 *** |
| 4.0 | 0.1 | 0.0 | 0.0 | 0.0 | 4.93 **** | 4.93 **** |
| 5.0 | 0.1 | 0.0 | 0.0 | 0.0 | 6.99 **** | 6.95 **** |
| | 1. | | | | | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

* THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 10 PERCENT.

** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 20 PERCENT.

*** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS GREATER THAN 20 PERCENT.

**** THE PLANT WOULD BE CURTAILED.

TABLE 9-37. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 5 ( 0.4064 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| | | 1980 $/MCF | | | | |
| 0.5 | 0.6 | 0.0 | 0.0 | 0.0 | 0.33 | 0.35 |
| 1.0 | 0.4 | 0.0 | 0.0 | 0.0 | 0.68 * | 0.41 * |
| 2.0 | 0.4 | 0.0 | 0.0 | 0.0 | 1.50 * | 1.77 * |
| 3.0 | 0.2 | 0.0 | 0.0 | 0.0 | 2.49 *** | 2.94 *** |
| 4.0 | 0.2 | 0.0 | 0.0 | 0.0 | 3.70 *** | 4.38 **** |
| 5.0 | 0.2 | 0.0 | 0.0 | 0.0 | 5.24 **** | 6.20 **** |
| | 2. | | | | | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

* THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 10 PERCENT.
** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 20 PERCENT.
*** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS GREATER THAN 20 PERCENT.
**** THE PLANT WOULD BE CURTAILED.

TABLE 9-38. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 6 ( 0.5679 M3/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE | | | | |
|---|---|---|---|---|---|---|
| | | 2 | 3 | 4 | 5 | 6 |
| | | 1980 $/MCF | | | | |
| 0.5 | 0.3 | 0.0 | 0.0 | 0.26 | 0.31 | 0.38 |
| 1.0 | 0.2 | 0.0 | 0.0 | 0.55 | 0.65 * | 0.79 * |
| 2.0 | 0.2 | 0.0 | 0.0 | 1.20 * | 1.42 * | 1.74 * |
| 3.0 | 0.1 | 0.0 | 0.0 | 1.95 ** | 2.35 ** | 2.89 *** |
| 4.0 | 0.1 | 0.0 | 0.0 | 2.96 *** | 3.50 *** | 4.50 **** |
| 5.0 | 0.1 | 0.0 | 0.0 | 4.15 **** | 4.96 **** | 5.09 **** |
| | 1. | | | | | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

* THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 10 PERCENT.

** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 20 PERCENT.

*** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS GREATER THAN 20 PERCENT.

**** THE PLANT WOULD BE CURTAILED.

TABLE 9-39. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 7 ( 0.6095 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE | | | | |
|---|---|---|---|---|---|---|
| | | 2 | 3 | 4 | 5 | 6 |
| | | 1980 $/MCF | | | | |
| 0.5 | 0.3 | 0.0 | 0.0 | 0.22 | 0.26 | 0.32 |
| 1.0 | 0.2 | 0.0 | 0.0 | 0.46 | 0.54 | 0.66 * |
| 2.0 | 0.2 | 0.0 | 0.0 | 1.00 * | 1.18 * | 1.45 * |
| 3.0 | 0.1 | 0.0 | 0.0 | 1.66 * | 1.96 ** | 2.41 *** |
| 4.0 | 0.1 | 0.0 | 0.0 | 2.47 *** | 2.92 *** | 3.58 *** |
| 5.0 | 0.1 | 0.0 | 0.0 | 3.49 *** | 4.14 **** | 5.08 **** |
| | 1. | | | | | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

*    THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 10 PERCENT.

**   THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 20 PERCENT.

***  THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS GREATER THAN 20 PERCENT.

**** THE PLANT WOULD BE CURTAILED.

TABLE 9-40. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 8 ( 0.7111 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| | | 1980 $/MCF | | | | |
| 0.5 | 0.3 | 0.0 | 0.0 | 0.19 | 0.22 | 0.27 |
| 1.0 | 0.2 | 0.0 | 0.0 | 0.35 | 0.46 | 0.57 |
| 2.0 | 0.2 | 0.0 | 0.0 | 0.86 * | 1.01 * | 1.24 * |
| 3.0 | 0.1 | 0.0 | 0.0 | 1.42 * | 1.68 * | 2.06 ** |
| 4.0 | 0.1 | 0.0 | 0.0 | 2.11 ** | 2.50 *** | 3.07 *** |
| 5.0 | 0.1 | 0.0 | 0.0 | 2.99 *** | 3.54 *** | 4.35 **** |
| | 1. | | | | | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

* THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 10 PERCENT.

** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 20 PERCENT.

*** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS GREATER THAN 20 PERCENT.

**** THE PLANT WOULD BE CURTAILED.

TABLE 9-41. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 9 ( 0.8127 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE | | | | |
|---|---|---|---|---|---|---|
| | | 2 | 3 | 4 | 5 | 6 |
| | | 1980 $/MCF | | | | |
| 0.5 | 0.3 | 0.0 | 0.0 | 0.16 | 0.19 | 0.24 |
| 1.0 | 0.2 | 0.0 | 0.0 | 0.34 | 0.40 | 0.50 |
| 2.0 | 0.2 | 0.0 | 0.0 | 0.75 * | 0.89 * | 1.09 * |
| 3.0 | 0.1 | 0.0 | 0.0 | 1.24 * | 1.47 * | 1.80 ** |
| 4.0 | 0.1 | 0.0 | 0.0 | 1.85 ** | 2.19 ** | 2.65 *** |
| 5.0 | 0.1 | 0.0 | 0.0 | 2.62 *** | 3.10 *** | 3.31 *** |
| | 1. | | | | | |

REGULATORY ALTERNATIVE I IS THE BASELINE.

* THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 10 PERCENT.

** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 20 PERCENT.

*** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS GREATER THAN 20 PERCENT.

TABLE 9-42. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 10 ( 0.9143 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| | | 1980 $/MCF | | | | |
| 0.5 | 0.3 | 0.0 | 0.0 | 0.15 | 0.17 | 0.21 |
| 1.0 | 0.2 | 0.0 | 0.0 | 0.30 | 0.36 | 0.44 |
| 2.0 | 0.2 | 0.0 | 0.0 | 0.67 * | 0.79 * | 0.97 * |
| 3.0 | 0.1 | 0.0 | 0.0 | 1.10 * | 1.31 * | 1.60 * |
| 4.0 | 0.1 | 0.0 | 0.0 | 1.64 * | 1.95 ** | 2.39 ** |
| 5.0 | 0.1 | 0.0 | 0.0 | 2.13 ** | 2.76 *** | 3.38 *** |
| | 1. | | | | | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

* THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 10 PERCENT.

** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 20 PERCENT.

*** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS GREATER THAN 20 PERCENT.

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TABLE 9-43. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT II (     1.0159 MG/CD)

| PERCENT $H_2S$ IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| | | 1980 $/MCF | | | | |
| 0.5 | 1.5 | 0.0 | 0.13 | 0.13 | 0.16 | 0.19 |
| 1.0 | 1.0 | 0.0 | 0.27 | 0.27 | 0.32 | 0.40 |
| 2.0 | 1.0 | 0.0 | 0.60 | 0.60 | 0.71 * | 0.87 * |
| 3.0 | 0.5 | 0.0 | 0.99 * | 0.99 * | 1.18 * | 1.44 * |
| 4.0 | 0.5 | 0.0 | 1.48 * | 1.48 * | 1.75 * | 2.15 ** |
| 5.0 | 0.5 | 0.0 | 2.10 ** | 2.10 ** | 2.48 *** | 3.05 *** |
| | 5. | | | | | |

REGULATORY ALTERNATIVE I IS THE BASELINE.
* THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 10 PERCENT.
** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 20 PERCENT.
*** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS GREATER THAN 20 PERCENT.

TABLE 9-44. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 12 ( 2.0313 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE<br>2 | <br>3 | <br>4 | <br>5 | <br>6 |
|---|---|---|---|---|---|---|
| | | 1980 $/MCF | | | | |
| 0.5 | 0.9 | 0.0 | 0.07 | 0.07 | 0.08 | 0.10 |
| 1.0 | 0.6 | 0.0 | 0.14 | 0.14 | 0.16 | 0.20 |
| 2.0 | 0.6 | 0.0 | 0.30 | 0.30 | 0.35 | 0.44 |
| 3.0 | 0.3 | 0.0 | 0.50 | 0.50 | 0.59 | 0.72 * |
| 4.0 | 0.3 | 0.0 | 0.74 * | 0.74 * | 0.89 * | 1.08 * |
| 5.0 | 0.3 | 0.0 | 1.05 * | 1.05 * | 1.24 * | 1.52 * |
| | 3. | | | | | |

REGULATORY ALTERNATIVE I IS THE BASELINE.

* THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 10 PERCENT.

TABLE 9-45. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 13 ( 2.0477 M3/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE | | | | |
|---|---|---|---|---|---|---|
| | | 2 | 3 | 4 | 5 | 6 |
| | | 1980 $/MCF | | | | |
| 0.5 | 0.9 | 0.0 | 0.04 | 0.05 | 0.05 | 0.06 |
| 1.0 | 0.6 | 0.0 | 0.09 | 0.11 | 0.11 | 0.13 |
| 2.0 | 0.6 | 0.0 | 0.20 | 0.24 | 0.24 | 0.25 |
| 3.0 | 0.3 | 0.0 | 0.33 | 0.39 | 0.39 | 0.48 |
| 4.0 | 0.3 | 0.0 | 0.49 | 0.58 | 0.58 | 0.72 * |
| 5.0 | 0.3 | 0.0 | 0.70 * | 0.83 * | 0.83 * | 1.02 * |
| | 3. | | | | | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

* THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 10 PERCENT.

TABLE 9-46. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 14 (    4.0636 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE | | | | |
|---|---|---|---|---|---|---|
| | | 2 | 3 | 4 | 5 | 6 |
| | | 1980 $/MCF | | | | |
| 0.5 | 0.6 | 0.0 | 0.03 | 0.04 | 0.04 | 0.05 |
| 1.0 | 0.4 | 0.0 | 0.07 | 0.08 | 0.08 | 0.10 |
| 2.0 | 0.4 | 0.0 | 0.15 | 0.18 | 0.18 | 0.22 |
| 3.0 | 0.2 | 0.0 | 0.25 | 0.29 | 0.29 | 0.36 |
| 4.0 | 0.2 | 0.0 | 0.37 | 0.44 | 0.44 | 0.54 |
| 5.0 | 0.2 | 0.0 | 0.52 | 0.62 * | 0.62 * | 0.76 * |
| | 2. | | | | | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

* THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 10 PERCENT.

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TABLE 9-47. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 15 ( 5.0795 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE | | | | |
|---|---|---|---|---|---|---|
| | | 2 | 3 | 4 | 5 | 6 |
| | | 1980 $/MCF | | | | |
| 0.5 | 0.3 | 0.0 | 0.03 | 0.03 | 0.03 | 0.04 |
| 1.0 | 0.2 | 0.0 | 0.05 | 0.06 | 0.06 | 0.08 |
| 2.0 | 0.2 | 0.0 | 0.12 | 0.14 | 0.14 | 0.17 |
| 3.0 | 0.1 | 0.0 | 0.20 | 0.24 | 0.24 | 0.27 |
| 4.0 | 0.1 | 0.0 | 0.30 | 0.35 | 0.35 | 0.43 |
| 5.0 | 0.1 | 0.0 | 0.42 | 0.50 | 0.50 | 0.61 * |
| 6.0 | 0.0 | 0.0 | 0.58 | 0.69 * | 0.69 * | 0.84 * |
| 7.0 | 0.0 | 0.0 | 0.80 * | 0.95 * | 0.95 * | 1.16 * |
| 8.0 | 0.0 | 0.0 | 1.12 * | 1.32 * | 1.32 * | 1.62 * |
| 9.0 | 0.0 | 0.0 | 1.61 * | 1.91 ** | 1.91 ** | 2.35 ** |
| 10.0 | 0.0 | 0.0 | 2.51 *** | 2.97 *** | 2.97 *** | 3.64 *** |
| | 1. | | | | | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

* THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 10 PERCENT.

** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 20 PERCENT.

*** THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS GREATER THAN 20 PERCENT.

TABLE 9-48. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 16 ( 10.1550 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| | | 1980 $/MCF | | | | |
| 0.5 | 2.0 | 0.0 | 0.00 | 0.00 | 0.01 | 0.01 |
| 1.0 | 2.0 | 0.0 | 0.00 | 0.00 | 0.02 | 0.02 |
| 2.0 | 2.0 | 0.0 | 0.01 | 0.01 | 0.05 | 0.05 |
| 3.0 | 1.0 | 0.0 | 0.01 | 0.01 | 0.08 | 0.08 |
| 4.0 | 1.0 | 0.0 | 0.01 | 0.01 | 0.10 | 0.10 |
| 5.0 | 1.0 | 0.0 | 0.01 | 0.01 | 0.13 | 0.13 |
| 6.0 | 0.0 | 0.0 | 0.02 | 0.02 | 0.16 | 0.16 |
| 7.0 | 0.0 | 0.0 | 0.02 | 0.02 | 0.19 | 0.19 |
| 8.0 | 0.0 | 0.0 | 0.03 | 0.03 | 0.23 | 0.23 |
| 9.0 | 0.0 | 0.0 | 0.03 | 0.03 | 0.26 | 0.26 |
| 10.0 | 0.0 | 0.0 | 0.03 | 0.03 | 0.30 | 0.30 |
| | 9. | | | | | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

TABLE 9-49. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 17 ( 10.1590 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE | | | | |
|---|---|---|---|---|---|---|
| | | 2 | 3 | 4 | 5 | 6 |
| | | 1980 $/MCF | | | | |
| 0.5 | 2.0 | 0.0 | 0.00 | 0.00 | 0.01 | 0.01 |
| 1.0 | 2.0 | 0.0 | 0.00 | 0.00 | 0.03 | 0.03 |
| 2.0 | 2.0 | 0.0 | 0.01 | 0.01 | 0.06 | 0.06 |
| 3.0 | 1.0 | 0.0 | 0.01 | 0.01 | 0.09 | 0.09 |
| 4.0 | 1.0 | 0.0 | 0.02 | 0.02 | 0.14 | 0.14 |
| 5.0 | 1.0 | 0.0 | 0.02 | 0.02 | 0.20 | 0.20 |
| 6.0 | 0.0 | 0.0 | 0.03 | 0.03 | 0.27 | 0.27 |
| 7.0 | 0.0 | 0.0 | 0.04 | 0.04 | 0.38 | 0.38 |
| 8.0 | 0.0 | 0.0 | 0.06 | 0.06 | 0.53 | 0.53 |
| 9.0 | 0.0 | 0.0 | 0.08 | 0.08 | 0.76 * | 0.76 * |
| 10.0 | 0.0 | 0.0 | 0.13 | 0.13 | 1.19 * | 1.19 * |
| | 9. | | | | | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

* THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 10 PERCENT.

TABLE 9-50. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 1B (    101.5900 MC/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE | | | | |
|---|---|---|---|---|---|---|
| | | 2 | 3 | 4 | 5 | 6 |
| | | 1980 $/MCF | | | | |
| 0.5 | 0.0 | 0.0 | 0.00 | 0.00 | 0.02 | 0.02 |
| 1.0 | 0.0 | 0.0 | 0.00 | 0.00 | 0.04 | 0.04 |
| 2.0 | 0.0 | 0.0 | 0.01 | 0.01 | 0.09 | 0.09 |
| 3.0 | 1.0 | 0.0 | 0.01 | 0.01 | 0.14 | 0.14 |
| 4.0 | 1.0 | 0.0 | 0.01 | 0.01 | 0.18 | 0.18 |
| 5.0 | 0.0 | 0.0 | 0.01 | 0.01 | 0.24 | 0.24 |
| 6.0 | 1.0 | 0.0 | 0.02 | 0.02 | 0.29 | 0.29 |
| 7.0 | 0.0 | 0.0 | 0.02 | 0.02 | 0.34 | 0.34 |
| 8.0 | 1.0 | 0.0 | 0.03 | 0.03 | 0.40 | 0.40 |
| 9.0 | 0.0 | 0.0 | 0.03 | 0.03 | 0.47 | 0.47 |
| 10.0 | 1.0 | 0.0 | 0.03 | 0.03 | 0.53 | 0.53 |
| | 5. | | | | | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

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TABLE 9-51. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT IV ( 101.5900 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| | | 1980 $/MCF | | | | |
| 0.5 | 0.0 | 0.0 | 0.00 | 0.00 | 0.02 | 0.02 |
| 1.0 | 0.0 | 0.0 | 0.00 | 0.00 | 0.04 | 0.04 |
| 2.0 | 0.0 | 0.0 | 0.01 | 0.01 | 0.09 | 0.09 |
| 3.0 | 0.0 | 0.0 | 0.01 | 0.01 | 0.15 | 0.15 |
| 4.0 | 1.0 | 0.0 | 0.01 | 0.01 | 0.21 | 0.21 |
| 5.0 | 0.0 | 0.0 | 0.02 | 0.02 | 0.28 | 0.28 |
| 6.0 | 1.0 | 0.0 | 0.02 | 0.02 | 0.36 | 0.36 |
| 7.0 | 0.0 | 0.0 | 0.03 | 0.03 | 0.46 | 0.46 |
| 8.0 | 1.0 | 0.0 | 0.04 | 0.04 | 0.57 | 0.57 |
| 9.0 | 0.0 | 0.0 | 0.04 | 0.04 | 0.69 * | 0.69 * |
| 10.0 | 1.0 | 0.0 | 0.05 | 0.05 | 0.85 * | 0.85 * |
| | 4. | | | | | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

* THE PROBABILITY THAT THE PLANT WOULD BECOME UNPROFITABLE IN THE LONG RUN IS APPROXIMATELY 10 PERCENT.

TABLE 9-52. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 20 ( 563.6242 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| | | 1980 $/MCF | | | | |
| 0.5 | 0.0 | 0.00 | 0.00 | 0.01 | 0.01 | 0.01 |
| 1.0 | 0.0 | 0.00 | 0.00 | 0.01 | 0.01 | 0.01 |
| 2.0 | 0.0 | 0.00 | 0.00 | 0.02 | 0.02 | 0.02 |
| 3.0 | 0.0 | 0.00 | 0.00 | 0.03 | 0.03 | 0.03 |
| 4.0 | 0.0 | 0.00 | 0.00 | 0.04 | 0.04 | 0.04 |
| 5.0 | 0.0 | 0.00 | 0.00 | 0.06 | 0.06 | 0.06 |
| 6.0 | 0.0 | 0.00 | 0.00 | 0.07 | 0.07 | 0.07 |
| 7.0 | 0.0 | 0.00 | 0.00 | 0.08 | 0.08 | 0.08 |
| 8.0 | 0.0 | 0.00 | 0.00 | 0.09 | 0.09 | 0.09 |
| 9.0 | 0.0 | 0.00 | 0.00 | 0.11 | 0.11 | 0.11 |
| 10.0 | 0.0 | 0.00 | 0.00 | 0.12 | 0.12 | 0.12 |
| 15.0 | 0.0 | 0.00 | 0.00 | 0.21 | 0.21 | 0.21 |
| 20.0 | 0.0 | 0.01 | 0.01 | 0.33 | 0.33 | 0.33 |
| | 0. | | | | | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

TABLE 9-53. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 21 ( 563.8242 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| | | 1980 $/MCF | | | | |
| 0.5 | 0.0 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| 1.0 | 0.0 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| 2.0 | 0.0 | 0.00 | 0.01 | 0.01 | 0.01 | 0.01 |
| 3.0 | 0.0 | 0.00 | 0.01 | 0.01 | 0.01 | 0.01 |
| 4.0 | 0.0 | 0.00 | 0.01 | 0.01 | 0.01 | 0.01 |
| 5.0 | 0.0 | 0.00 | 0.02 | 0.02 | 0.02 | 0.02 |
| 6.0 | 0.2 | 0.00 | 0.02 | 0.02 | 0.02 | 0.02 |
| 7.0 | 0.0 | 0.00 | 0.03 | 0.03 | 0.03 | 0.03 |
| 8.0 | 0.2 | 0.00 | 0.03 | 0.03 | 0.03 | 0.03 |
| 9.0 | 0.0 | 0.00 | 0.04 | 0.04 | 0.04 | 0.04 |
| 10.0 | 0.2 | 0.00 | 0.04 | 0.04 | 0.04 | 0.04 |
| 15.0 | 0.4 | 0.00 | 0.07 | 0.07 | 0.07 | 0.07 |
| 20.0 | 1.0 | 0.00 | 0.09 | 0.09 | 0.09 | 0.09 |
| | 2. | | | | | |

REGULATORY ALTERNATIVE 1 IS THE BASELINE.

TABLE 9-54. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 22 ( 1015.8997 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| | | 1980 $/MCF | | | | |
| 0.5 | 0.0 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| 1.0 | 0.0 | 0.00 | 0.00 | 0.01 | 0.01 | 0.01 |
| 2.0 | 0.0 | 0.00 | 0.00 | 0.02 | 0.02 | 0.02 |
| 3.0 | 0.0 | 0.00 | 0.00 | 0.03 | 0.03 | 0.03 |
| 4.0 | 0.0 | 0.00 | 0.00 | 0.04 | 0.04 | 0.04 |
| 5.0 | 0.0 | 0.00 | 0.00 | 0.05 | 0.05 | 0.05 |
| 6.0 | 0.0 | 0.00 | 0.00 | 0.06 | 0.06 | 0.06 |
| 7.0 | 0.0 | 0.00 | 0.00 | 0.07 | 0.07 | 0.07 |
| 8.0 | 0.0 | 0.00 | 0.00 | 0.08 | 0.08 | 0.08 |
| 9.0 | 0.0 | 0.00 | 0.00 | 0.10 | 0.10 | 0.10 |
| 10.0 | 0.0 | 0.00 | 0.00 | 0.11 | 0.11 | 0.11 |
| 15.0 | 0.0 | 0.00 | 0.00 | 0.19 | 0.19 | 0.19 |
| 20.0 | 0.0 | 0.00 | 0.00 | 0.29 | 0.29 | 0.29 |
| | 0. | | | | | |

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TABLE 9-55. MODEL PLANT EMISSION CONTROL COST PER MCF. (MODEL PLANT 23 ( 1015.8997 MG/D))

| PERCENT H2S IN SOUR NATURAL GAS | ASSUMED PLANT DISTRIBUTION | REGULATORY ALTERNATIVE 2 | 3 | 4 | 5 | 6 |
|---|---|---|---|---|---|---|
| | | 1980 $/MCF | | | | |
| 0.5 | 0.0 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| 1.0 | 0.0 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| 2.0 | 0.0 | 0.00 | 0.01 | 0.01 | 0.01 | 0.01 |
| 3.0 | 0.0 | 0.00 | 0.01 | 0.01 | 0.01 | 0.01 |
| 4.0 | 0.0 | 0.00 | 0.01 | 0.01 | 0.01 | 0.01 |
| 5.0 | 0.0 | 0.00 | 0.01 | 0.01 | 0.01 | 0.01 |
| 6.0 | 0.1 | 0.00 | 0.02 | 0.02 | 0.02 | 0.02 |
| 7.0 | 0.0 | 0.00 | 0.02 | 0.02 | 0.02 | 0.02 |
| 8.0 | 0.1 | 0.00 | 0.02 | 0.02 | 0.02 | 0.02 |
| 9.0 | 0.0 | 0.00 | 0.03 | 0.03 | 0.03 | 0.03 |
| 10.0 | 0.1 | 0.00 | 0.03 | 0.03 | 0.03 | 0.03 |
| 15.0 | 0.2 | 0.00 | 0.05 | 0.05 | 0.05 | 0.05 |
| 20.0 | 0.5 | 0.00 | 0.07 | 0.07 | 0.07 | 0.07 |
| | 1. | | | | | |

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control costs exceeds $1.80 per Mcf, they are marked with two asterisks (**) indicating a 20 percent estimated probability that any given model plant would cease to cover total production costs due to the regulatory alternative. Three asterisks (***) indicate that the incremental control costs are over $2.40 per Mcf, a level which exceeds half of the projected baseline price for new natural gas in 1987 and which would indicate more than a 20 percent probability that any given model plant would cease to cover total production costs due to the regulatory alternative. As shown in Table 9-33, four asterisks (****) indicate that unit control costs exceed $3.87 per Mcf, a level certain to cause curtailments since average variable operating costs would exceed the price of natural gas.

No plant curtailments are predicted under Regulatory Alternatives I-IV. Under Regulatory Alternative V, one plant would likely be curtailed. Under Regulatory Alternative VI, two plants would likely be curtailed. The decline in natural gas production associated with these potential curtailments would be negligible.

The above projection of curtailments is based on the assumption that exploration and development of new gas fields would be unaffected by the regulation. In order to qualitatively assess the reasonableness of this assumption, the increase in the number of plants that fail to cover total production costs is calculated for each regulatory alternative utilizing the methodology described in Section 9.2.1.2.

The results of this analysis indicate that, under Regulatory Alternatives I-III, costs are not expected to cause any increase in the number of plants unable to cover total production costs. Therefore, these three regulatory alternatives would have virtually no effect on exploration and development of new natural gas fields.

At most, nine small plants are estimated to become unprofitable in the long run under Regulatory Alternative VI. To obtain a more direct measure of the impact of Regulatory Alternatives I through VI on expected returns to natural gas exploration, the ratio of the expected aggregate cost of each regulatory alternative to the wellhead value of projected new gas production is calculated. These estimates are presented below:

| Regulatory alternative | Estimated aggregate after-tax annualized cost (millions 1980$) | Projected wellhead value of new onshore natural gas production 1/ (millions of 1980$) | Cost to value ratio (percent) |
|---|---|---|---|
| I | 0.0 | 11,600 | 0.0 |
| II | 0.0 | 11,600 | 0.0 |
| III | 12.2 | 11,600 | 0.1 |
| IV | 13.5 | 11,600 | 0.1 |
| V | 51.9 | 11,600 | 0.4 |
| VI | 53.5 | 11,600 | 0.5 |

Under Regulatory Alternatives I through VI, the aggregate costs of compliance are less than 1 percent of the total projected value of new onshore natural gas production in 1987. Thus, under Regulatory Alternatives I through VI, the impacts of $SO_2$ emissions control costs on expected returns from natural gas exploration and development are relatively small. These results suggest that the effects of Regulatory Alternatives I through VI on exploration and development are likely to be negligible.

9.2.2.3 Expected Quantity and Price Impacts Due to $SO_2$ NSPS. Sulfur dioxide emissions regulations on new onshore sour natural gas production facilities are not expected to cause significant quantity or price impacts in either the natural gas or sulfur markets. Changes in natural gas production or price are negligible under Regulatory Alternatives I through VI since the number of curtailed plants is zero or negligible.

Sulfur recovery from onshore natural gas would increase slightly as a result of regulations that require a greater percentage of available sulfur to be recovered. This would increase new onshore natural gas sulfur recovery from 1.2 million metric tons in the baseline to 1.3 million metric tons under Regulatory Alternatives IV, V and VI. This represents about a 0.5 percent increase in total projected sulfur supply in 1987. This could cause a slight decrease in local sulfur prices in markets near new natural gas sulfur recovery. These effects would most likely occur in Texas, New Mexico, Wyoming, Alabama, Mississippi, Arkansas, and North Dakota where new onshore sour natural gas discoveries can be expected.

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1/ Based on 2.42 trillion cubic feet (Table 9-20) and $4.80/Mcf.

## 9.3 POTENTIAL SOCIOECONOMIC AND INFLATIONARY IMPACTS

Sulfur dioxide emissions regulations on new onshore sour natural gas sulfur recovery facilities are not expected to cause major socioeconomic or inflationary impacts. Curtailments in the construction of new plants are projected to be negligible.

Total additional before-tax annualized costs of controls in 1987, the fifth year of controls, for the projected sixty-seven new natural gas sulfur recovery plants are estimated to be as follows:

| Regulatory alternative, $SO_2$ | Total additional before-tax annualized cost, 1987 [a] (million 1980 dollars) |
|---|---|
| I | 0.0 |
| II | 0.0 |
| III | 22.2 |
| IV | 24.7 |
| V | 95.3 |
| VI | 98.7 |

---

[a] These estimates are obtained from Table 9-56.

Individual natural gas sulfur recovery plant operators are not expected to be able to pass forward additional sulfur emissions control costs to consumer sectors. Costs may be passed backward to sour gas producers, but often natural gas sulfur recovery and sour natural gas production will be vertically integrated. Thus, sour gas producers are generally expected to absorb the additional emissions control costs.

A slight increase in sulfur production may occur as a result of increased sulfur recovery. This may cause sulfur prices to fall slightly. This impact, however, is expected to be small and regional in nature. Sulfur consuming industries, particularly fertilizer manufacturers, could benefit from slightly expanded sulfur supply and lower prices. Sulfur importers to the U.S. and the Frasch mining segment of the sulfur industry may be adversely impacted by increased sour natural gas sulfur recovery, but again expected impacts will be slight regardless of the regulatory alternative chosen.

### Table 9-56. TOTAL BEFORE TAX NET ANNUALIZED COST [a] REGULATORY ALTERNATIVES, 1987

| Model plant | Size (sulfur intake) Mg/d | Size (sulfur intake) LT/d | Number of model plants | Regulatory alternative II | III | IV | V | VI |
|---|---|---|---|---|---|---|---|---|
| | | | | Millions of 1980 dollars | | | | |
| 1 | <0.1 | <0.1 | 11 | 0 | 0 | 0 | 0 | 0 |
| 2 | 0.1 | 0.1 | 1 | 0 | 0 | 0 | 0 | 0 |
| 3 | 0.2 | 0.2 | 3 | 0 | 0 | 0 | 0 | 1.3 |
| 4 | 0.3 | 0.3 | 1 | 0 | 0 | 0 | .4 | .4 |
| 5 | 0.4 | 0.4 | 2 | 0 | 0 | 0 | .8 | 1.0 |
| 6 | 0.5 | 0.5 | 1 | 0 | 0 | .4 | .5 | .6 |
| 7 | 0.6 | 0.6 | 1 | 0 | 0 | .4 | .5 | .6 |
| 8 | 0.7 | 0.7 | 1 | 0 | 0 | .4 | .5 | .6 |
| 9 | 0.8 | 0.8 | 1 | 0 | 0 | .4 | .5 | .6 |
| 10 | 0.9 | 0.9 | 1 | 0 | 0 | .4 | .5 | .6 |
| 11 | 1 | 1 | 5 | 0 | 2.1 | 2.1 | 2.5 | 3.0 |
| 12 | 2 | 2 | 3 | 0 | 1.3 | 1.3 | 1.5 | 1.8 |
| 13 | 3 | 3 | 3 | 0 | 1.3 | 1.5 | 1.5 | 1.8 |
| 14 | 4 | 4 | 2 | 0 | .8 | 1.0 | 1.0 | 1.2 |
| 15 | 5.1 | 5 | 1 | 0 | .4 | .5 | .5 | .6 |
| 16-17 | 10.2 | 10 | 18 | 0 | .8 | .8 | 7.0 | 7.0 |
| 18-19 | 101.6 | 100 | 9 | 0 | 4.1 | 4.1 | 66.2 | 66.2 |
| 20 | 563.8 | 555 | 0 | 0 | 0 | 0 | 0 | 0 |
| 21 | 563.8 | 555 | 2 | -.1 | 6.7 | 6.7 | 6.7 | 6.7 |
| 22 | 1,015.9 | 1,000 | 0 | 0 | 0 | 0 | 0 | 0 |
| 23 | 1,015.9 | 1,000 | 1 | -.007 | 4.7 | 4.7 | 4.7 | 4.7 |
| Total | | | 67 | -.107 | 22.2 | 24.7 | 95.3 | 98.7 |

Regulatory Alternative I is the baseline.

[a] The total costs for each model plant segment equals the number of model plants per segment times the before-tax annualized cost per plant. (Table 9-31)

No significant net employment, productivity or balance of payments impacts are expected as a result of sulfur emissions regulations on the onshore natural gas production industry. Some slight displacements could occur if the natural gas industry shifts production resources away from known marginal sour gas reserves and towards exploration and development of known sweet gas reserves.

## 9.4 REFERENCES FOR CHAPTER 9

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3. American Petroleum Institute, 1979 Joint Association Survey on Drilling Costs, February 1981. Docket No. A-80-20-A, Entry II-I-77.

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13. A. Bernhard & Company. "Natural Gas Industry, Value Line Investment Survey," July 18, 1980. Docket No. A-80-20-A, Entry II-I-79.

14. Stanford Research Institute (SRI) International, Chemical Economics Handbook, December 1979. Docket No. A-80-20-A, Entry II-I-44.

15. American Gas Association, "Total Energy Resource Analysis Model (TERA) 80-1," Gas Supply and Statistics, Appendix A, Figure A-2, p. 21, 1980. Docket No. A-80-20-A, Entry II-I-62.

16. U.S. Department of Energy, Energy Information Administration, DOE/EIA-0173(79)/3, Annual Report to Congress - 1979, Volume Three (of three): Projections. Docket No. A-80-20-A, Entry II-I-46.

17. TRW Environmental Division, Research Triangle Park, North Carolina, for U.S. EPA, Office of Air Quality Planning and Standards, Emission Standards and Engineering Division, Source Category Survey Report - Phase I Onshore Production, March 19, 1980. Docket No. A-80-20-A, Entry II-A-13.

18. Personal communication between Mr. K. Joshi of TRW Environmental Division with Mr. Richard M. Schulze of Trinity Consultants in Richardson, Texas, on April 23, 1981. Docket No. A-80-20-A, Entry II-E-16.

19. Documented meeting between TRW, EPA and API Task Force held July 21 and 22, 1980 at TRW Environmental Division offices in Research Triangle Park, North Carolina. Docket No. A-80-20-A, Entry II-E-6.

20. U.S. Environmental Protection Agency, Industrial Environmental Research Laboratory, Research Triangle Park, North Carolina, Multimedia Assessment of the Natural Gas Processing Industry, EPA-600/2-79-077, April 1979, Section 5, p. 28. Docket No. A-80-20-A, Entry II-A-12.

21. U.S. Department of Energy, Energy Information Administration, DOE/EIA-0173(79)/3, Annual Report to Congress - 1979, Volume Three (of three): Projections. Docket No. A-80-20-A, Entry II-I-46.

22. Personal communication between Mr. K. Joshi of TRW Environmental Division with Mr. E. E. Ellington of Purwin & Gertz, Inc. in Dallas, Texas, on August 27, 1981. Docket No. A-80-20-A, Entry II-E-23.

23. Personal communication between Mr. K. Joshi of TRW Environmental Division with Mr. James Myers of Texas Air Control Board, Permit Section in Austin, Texas, on August 27, 1981. Docket No. A-80-20-A, Entry II-E-24.

24. Personal communication between Mr. K. Joshi of TRW Environmental Division with Mr. David Parnell of Ford, Bacon and Davis Engineers and Constructors in Dallas, Texas, on August 28, 1981. Docket No. A-80-20-A, Entry II-E-25.

25. Data Resources, Inc., Trendlong 2005 Forecasts, September 1980. Docket No. A-80-20-A, Entry II-I-81.

26. U.S. Energy Outlook - Oil and Gas Availability, National Petroleum Council, 1974. Docket No. A-80-20-A, Entry II-I-82.

# APPENDIX A - EVOLUTION OF THE BACKGROUND INFORMATION DOCUMENT

## A.1 LITERATURE REVIEW

| Date | Source | Data or Information |
|---|---|---|
| August 28, 1974 | (Report)<br>EPA-68-02-0611 | Characterization of Sulfur Recovery in Oil and Natural Gas Production |
| December 1974 | (Report)<br>EPA-650/2-75-030 | Sulfur Compound Emissions of the Petroleum Production Industry |
| October 1975 | (Report)<br>EPA-450/3-75-076 | Atmospheric Emissions Survey of the Sour Gas Processing Industry |
| February 20, 1976 | (Report)<br>EPA-68-02-2082 | Economic Impact of New Source Performance Standards on Sulfur Recovery Plants Associated with Natural Gas Processing Plants |
| January 1977 | (Draft report)<br>EPA | An Investigation of the Best Systems of Emissions Reduction for Sulfur Compounds from Crude Oil and Natural Gas Processing Plants |
| March 2 and 3, 1977 | U.S. EPA NAPCTAC Minutes of Meeting (first NAPCTAC) | Comments from API and the industry representatives on the proposed $SO_2$ NSPS on crude oil and natural gas onshore production plants. |
| August 1977 | (Report) EPA<br>AP-42 | Compilation of Air Pollutant Emission Factors, Third Edition |
| August 1978 | (Report)<br>EPA-450/3-78-047 | Evaluation of Emissions from Onshore Drilling, Producing, and Storing of Oil and Gas |
| January 8, 1980 | Rockwell International | Offshore program data |
| February 7, 1980 | Shell Oil Company<br>Houston, Texas | Forecast data for the lower 48 States' onshore natural gas production conventional facilities |

| Date | Source | Data or Information |
|---|---|---|
| July 29, 1980 | Energy Resources Conservation Board, the Province of Alberta, Canada | Sour Natural Gas Industry Guidelines (Existing) |
| October 6, 1980 | Energy Resources Conservation Board, the Province of Alberta, Canada | Sour Natural Gas Industry Guidelines (newly promulgated) |
| November 25, 1980 | Benfield Corporation Pittsburgh, PA | Cost information for the Benfield Sweetening Process for sour natural gas |
| February 20, 1981 | Getty Oil Company Houston, Texas | Comments and recommendations on sulfur recovery operations |
| April 14, 1981 | Home Oil Company, Ltd. Alberta, Canada | Technical and cost data on the modified Claus sulfur recovery facility for Carstairs-Crossfield plant |
| April 29, 1981 | Ralph M. Parsons Co. Pasadena, California | Information and assumptions for developing design criteria and cost estimates for sulfur recovery study in onshore sour gas production facilities, presented in Appendix E |
| May 12, 1981 | Lone Star Gas Company Warwink, Texas | Recovered sulfur production from sour natural gas stream (acid gas) at Warwink Gas Treating Plant |
| May 27, 1981 | Northern Natural Gas Company, Midland, Texas | Recovered sulfur production from sour natural gas stream (acid gas) at Hobbs Sulfur Recovery facility |
| July 1, 1981 | Colorado Interstate Gas Company, Table Rock, Wyoming | Recovered sulfur information |
| July 1981 | (Subcontracted report) The Ralph M. Parsons Co. Pasadena, California | Sulfur Recovery Study, Onshore Sour Gas Production Facilities, presented in Appendix E |
| September 3, 1981 | Exxon Chemical Co. Baton Rouge, LA | EPA requested costs information, including total operating costs and fixed-capital costs of the Baytown, Texas, refinery sulfur recovery plant by authority under Section 114 of the Clean Air Act (42 U.S.C. 7414) |

| Date | Source | Data or Information |
|---|---|---|
| September 3, 1981 | Pursue Gas Processing and Petrochemicals Co.<br>Jackson, Mississippi | EPA requested costs information, including total operating costs and fixed-capital costs of the Brandon, Mississippi, sour natural gas production sulfur recovery plant by authority under Section 114 of the Clean Air Act (42 U.S.C. 7414) |
| September 3, 1981 | Union Oil Company<br>Wilmington, California | EPA requested costs information, including total operating costs and fixed-capital costs of the Los Angeles, California, refinery sulfur recovery plant by authority under Section 114 of the Clean Air Act (42 U.S.C. 7414) |
| July 1982 | The American Petroleum Institute (API) | The API's gas plant survey containing data/information on 731 gas streams. The data included July 1982 capacity, throughput and gas stream $H_2S$ and $CO_2$ concentrations for onshore sour natural gas sweetening and sulfur recovery operations. |
| April 1983 | (Subcontract report)<br>The Ralph M. Parsons Co.<br>Pasadena, California | Small sulfur recovery (Recycle Selectox process) units, onshore sour gas production facilities, presented in Appendix H |

## A.2 PLANT VISITS

| Date | Company | Plant/Information |
|---|---|---|
| December 18, 1979 | Exxon Company, U.S.A.<br>Jay Field, Florida | Blackjack Creek sulfur recovery facility/gained familiarity with process equipment and operating conditions (sour natural gas sweetening, sulfur recovery, and tail gas cleanup operations), information confidential |
| December 19, 1979 | Phillips Petroleum<br>Chatham, Mississippi | Chatham sulfur recovery facility/gained familiarity with process equipment and operating conditions (sour natural gas sweetening and sulfur recovery operations) |

| Date | Source | Data or Information |
|---|---|---|
| August 8, 1980 | Shell Canada Resources, Ltd.<br>Waterton, Alberta, Canada | Waterton sulfur recovery facility/gained familiarity with process equipment and operating conditions (sour natural gas sweetening, sulfur recovery, and tail gas cleanup operations) |
| August 15, 1980 | Union Oil Company<br>Chunchula, Alabama | Chunchula sulfur recovery facility/gained familiarity with process equipment and operating conditions (sour natural gas sweetening and sulfur recovery operations) |
| August 19, 1980 | Getty Oil Company<br>Streetman, Texas | Teas sulfur recovery facility/ gained familiarity with process equipment and operating conditions (sour natural gas sweetening and sulfur recovery operations) |
| August 21, 1980 | Intratex Gas Company, (Houston Natural Gas), Pecos, Texas | Mi Vida sulfur recovery facilities/ gained familiarity with process equipment and operating conditions (sour natural gas sweetening and sulfur recovery operations), information confidential |
| August 22, 1980 | Warren Petroleum Company, (Gulf Oil), Monument, New Mexico | Monument sulfur recovery facility/ gained familiarity with process equipment and operating conditions (sour natural gas sweetening and sulfur recovery operations), information confidential |
| August 26, 1980 | Shell Oil Company<br>Brandon, Mississippi | Thomasville sulfur recovery facility/gained familiarity with process equipment and operating conditions (sour natural gas sweetening and sulfur recovery operations) |
| September 9, 1980 | Warren Petroleum Company, (Gulf Oil), Kildeer, North Dakota | Little Knife sulfur recovery facility/gained familiarity with process equipment and operating conditions (sour natural gas sweetening, sulfur recovery, and tail gas cleanup operations), information confidential |

| Date | Source | Data or Information |
|---|---|---|
| August 12, 1982 | Perry Gas Processors, Odessa, Texas | Perry gas processing facility/ gained familiarity with process and operating data on their Recycle Selectox process. A part of the information confidential. |
| August 12, 1982 | Sid Richardson Carbon and Gasoline Company, Winkler County, near Kermit, Texas | Sid Richardson Carbon and Gasoline's Recycle Selectox 3-stage sulfur recovery facility. Gained familiarity with process equipment, and operating conditions on their Recycle Selectox facility that recovers liquid sulfur from lean acid gas stream. Information confidential. |

A.3 EMISSION SOURCE TESTING

| Date | Company | Data or Information |
|---|---|---|
| March 5-27, 1981 | Warren Petroleum's Monument Plant Facility, Monument, New Mexico | Emission rates of $SO_2$, $H_2S$, total reduced sulfur, and $NO_x$; liquid sulfur production rate; sulfur recovery efficiency, plant operating conditions. Developed by and received from EPA/EMB. |
| April 8-12, 1981 | Getty Oil's New Hope Facility, Mt. Pleasant Texas | Emission rates of $SO_2$, $H_2S$, total reduced sulfur, and $NO_x$; liquid sulfur production rate; sulfur recovery efficiency, plant operating conditions. Developed by and received from EPA/EMB. |
| May 12-18, 1981 | Shell Oil's Thomasville Facility, Brandon, Mississippi | Emission rates of SO, $H_2S$, total reduced sulfur, and $NO_x$; liquid sulfur production rate; sulfur recovery efficiency, plant operating conditions. Developed by and received from EPA/EMB. |
| 1980, 1979, 1978, and 1977 | Exxon Company, U.S.A. Jay Field, Florida | Annual stack test reports (last four years) on Blackjack Creek sulfur recovery facility. Supplied by the facility. |

| Date | Company | Plant/Information |
|---|---|---|
| 1975, 1974, 1973, and 1972 | Shell Oil Company<br>Cass County, Texas | General information, flow diagrams, and operation and test data on Bryans Mill plant sulfur recovery facility. Supplied by the facility. |
| 1975, 1974 and 1973 | Shell Oil Company<br>Karnes County, Texas | General information, flow diagrams, and operation and test data on Person plant sulfur recovery facility. Supplied by the facility. |
| June 4-10, 1975 | Exxon Company, U.S.A.<br>Jay Field, Florida | Air pollution emission test, composition and flow data on Santa Rosa plant sulfur recovery facility. Developed by and received from EPA. |
| November 12, 1973 | Exxon Company, U.S.A.<br>Flomaton, Alabama | Performance test, composition and flow data on Flomaton sulfur recovery facility. Developed by and received from EPA. |
| September 24-27, 1974 | Aquitaine of Canada, Ltd.<br>Ram River, Alberta, Canada | Air pollution emission test on Ram River plant sulfur recovery facility. Developed by and received from EPA. |
| November 6-8, 1974 | Chevron Standard, Ltd.<br>Fox Creek, Alberta, Canada | Emission tests on Fox Creek plant sulfur recovery facility. Developed by and received from EPA. |

## A.4 MEETINGS WITH INDUSTRY

| Date | Attendees | Topic |
|---|---|---|
| December 1979 | American Petroleum Institute (Environmental Task Force) | The natural gas production industry |
| July 1980 | American Institute of Chemical Engineers, Philadelphia, PA | Seminars concerning sour natural gas processing technologies |
| July 21-22, 1980 | American Petroleum Institute (Environmental Task Force) and other representatives of the natural gas production industry | Exchange of information among TRW, EPA and the industry on various agenda concerning natural gas production |

| Date | Attendees | Topic |
|---|---|---|
| May 1, 1981 | American Petroleum Institute (Environmental Task Force) | Exchange of information among TRW, EPA and the industry on model plants, Parsons' cost estimations, and regulation format, etc. |
| January 28, 1982 | American Petroleum Institute (Environmental Task Force) | A presentation on cost effectiveness of the Claus 2-stage technology when applied to sizes less than 10.2 Mg/d sulfur intake. A presentation on projected growth in the industry for 1983-1987. Discussions among TRW, EPA and the industry on the items in the Agenda involving both $SO_2$ and VOC NSPS developments. |
| July 13, 1982 | The Ralph M. Parsons Co., Pasadena, California, and the Union Oil Co. of California, Brea, California | A presentation by the Parsons and Union Oil on the Recycle Selectox 2-stage and 3-stage processes to recover liquid sulfur from lean acid gas streams. Discussions among TRW, EPA and the Parsons on the methodology used by the Parsons to estimate catalyst performance degradation. |
| December 1, 1982 | Texas Eastern Gas Pipeline Co., Houston, Texas | Discussions of Texas Eastern Gas Pipeline Company's comments on the draft proposed NSPS for $SO_2$ emissions in the natural gas production industry with enclosed information on their analysis on incremental cost impact. |
| January 13, 1983 | Texas Oil and Gas Corporation, Dallas, Texas and the Independent Petroleum Association of America, Washington, D.C. | Discussions of Texas Oil and Gas Corporation's comments on the small plant size cutoff and the potential economic impacts from regulating small plants. |

| Date | Attendees | Topic |
|---|---|---|
| July 19, 1983 | Union Oil Company, Brea, CA and the Ralph M. Parsons Co. | Discussed response to EPA/TRW questions on the capabilities and limitations of the Recycle Selectox process and the Selectox® catalyst. |

## A.5 REVIEW PROCESS

| Date | Data or Information |
|---|---|
| August 14, 1981 | EPA concurs on the baseline control levels, the form/number of model plants and the regulatory alternatives |
| September 25, 1981 | Draft Chapters 3, 4, 5, and 6 mailed to the industry. |
| September 25, 1981 | Finalized concurrence memo on baseline controls, seven model plants and six regulatory alternatives. |
| October 30, 1981 | Submittal of tabular costs (Tables 8-1 through 8-42) to EPA to be included in Chapter 8 of the Background Information Document; (cost effectiveness, incremental cost effectiveness of the six regulatory alternatives for each of the seven model plants, total annualized costs, sulfur and steam credits, net annualized costs per megagram recovered sulfur, fixed-capital costs, sulfur recovery efficiency and $SO_2$ emissions, and sweetening operations costs for all the model plants) |
| December 23, 1981 | Decision on basis for the standard recommending NSPS development, including recommended size cutoffs for affected facilities. |
| February 1982 | Analyzed industry comments on Draft Chapters 3-6 of the Background Information Document |

| Date | Data or Information |
|---|---|
| April 9, 1982 | Draft copies of the Background Information Document, Preamble, and Regulation submitted to EPA for the EPA's Working Group review. |
| July 21, 1982 | EPA presentation to NAPCTAC Committee of the preliminary results of their development of a new source performance standard for $SO_2$ emissions from the Natural Gas Production Industry. The meeting was held in Dallas, Texas. |
| August 23, 1982 | Minutes of the NAPCTAC meeting held in Dallas, Texas on July 21, 1982 were distributed to the NAPCTAC Committee members, Industry representatives and various Interest Groups. |
| January, 1983 | Draft copies of the Background Information Document, Preamble and Regulation submitted to EPA for AA concurrence package for proposal in *Federal Register*. |
| June, 1983 | Revised draft copies of the background information document, preamble and regulation submitted to EPA for AA concurrence. Revisions incorporate new cost and performance data for Recycle Selectox technologies on small plants. |

# APPENDIX B - INDEX TO ENVIRONMENTAL IMPACT CONSIDERATIONS

Table B-1. INDEX TO ENVIRONMENTAL IMPACT CONSIDERATIONS

| Agency Guidelines for Preparing Regulatory Action Environmental Impact Statements (39 FR 37419) | Location Within the Background Information Document (BID) |
|---|---|
| 1. Background, description, and purpose of the regulatory alternatives and the statutory authority. | The regulatory alternatives from which standards will be chosen are summarized in Chapter 1, Section 1.1 and Chapter 6, Section 6.2, as is the statutory authority for proposing standards. |
| The relationship to other actions and proposals significantly affected by the regulatory alternatives. | To the extent possible, other EPA regulations and OSHA regulations which apply to the affected industries are detailed in Chapter 8, Section 8.1 and are considered in the economic impact study in Chapter 9. |
| Industry affected by the regulatory alternatives. | The industry and emission sources within the industry affected by the regulatory alternatives are listed in Chapter 3. |
| Specific processes affected by the regulatory alternatives. | The specific processes and facilities affected by the regulatory alternatives are summarized in Chapter 3, Section 3.2. |
| Applicable control technologies. | The sulfur recovery technologies which can be applied to reduce sulfur emissions from onshore natural gas processing facilities and the emission reductions which have been achieved by these control technologies are described in detail in Chapter 4. |
| 2. Alternatives to the action. | The various categories of alternatives to the actions which were considered are listed below.<br>a. Alternative regulatory approaches. The alternative approaches for regulating $SO_2$ emissions under Section 111 of the Clean Air Act are outlined in Chapter 6.<br>b. Alternative control devices. The alternative control devices that could be utilized by the regulatory alternatives, and the reasons for selecting these particular alternatives are outlined in Chapter 4. |

(continued)

Table B-1. Continued

| Agency Guidelines for Preparing Regulatory Action Environmental Impact Statements (39 FR 37419) | Location Within the Background Information Document (BID) |
|---|---|
| Agency's comparative evaluation of the beneficial and adverse environmental, health, social, and economic effects of each reasonable alternative. | a. A discussion of the Agency's comparative evaluation of the various alternative regulatory approaches for sulfur dioxide emissions from onshore natural gas processing facilities can be found in Chapter 6, Section 6.2. |
| | b. A summary of the beneficial and adverse environmental effects of the regulatory alternatives can be found in Chapter 7. A detailed description of the economic impacts of each alternative control level, including the capital and annualized costs to the industry, can be found in Chapter 8. The socioeconomic impacts of the regulatory alternatives including potential plant closures and maximum price increases in sweetened natural gas can be found in Chapter 9. |
| 3. Environmental impact of the regulatory alternatives. | |
| a. Primary impact. | |
| Primary impacts are those that can be attributed directly to the action, such as reduced levels of specific pollutants brought about by a new standard and the physical changes that occur in the various media with this reduction. | Air impact analysis shows the primary impacts in Chapter 7, Section 7.1.2. |

(continued)

Table B-1. Concluded

| Agency Guidelines for Preparing Regulatory Action Environmental Impact Statements (39 FR 37419) | Location Within the Background Information Document (BID) |
|---|---|
| b. Secondary impact. | |
| Secondary impacts are indirect or induced impacts. For example, mandatory reduction of specific pollutants brought about by a new standard could result in the adoption of control technology that exacerbates another pollution problem and would be a secondary impact. | Other environmental impacts of the individual controls that can be used to meet the regulatory alternatives are identified qualitatively in Chapter 7, Sections 7.1.3, 7.2, and 7.3.<br>The secondary water impacts of the alternative control levels are quantified in Chapter 7, Section 7.2.<br>The energy consumption impact of the alternative control levels is quantified in Chapter 7, Section 7.4. |
| 4. Other considerations. | |
| a. Adverse impacts which cannot be avoided should a regulatory alternative be implemented. | A summary of the potential adverse environmental and health impacts of the regulatory alternatives and a discussion of the significance of each impact can be found in Chapter 7. Factors which already exist to eliminate some of the potential adverse impacts are identified. Those adverse impacts that are unavoidable are also identified, along with any steps which can be taken to minimize them. |
| b. Irreversible and irretrievable commitments of resources that would be involved with the regulatory alternatives, should one be implemented. | Irreversible and irretrievable resources that would be involved in the proposed action are discussed in Chapter 7, Section 7.5.1. |

# APPENDIX C - EMISSION SOURCE TESTS DATA

The emission source tests data on ten Claus sulfur recovery facilities gathered during the development of the proposed Natural Gas Production New Source Performance Standards are summarized in this appendix. As part of the data gathering process, detailed emission source tests were conducted by EPA's Emissions Measurement Branch at three facilities. These facilities are Warren Petroleum's Monument plant facility, Getty Oil's New Hope facility and Shell Oil's Thomasville facility and are described in Section C.1. These facilities represent the existing industry practice for Claus sulfur recovery technology, for the size range of sulfur intake (sulfur feed rate) to sulfur recovery unit, and for the hydrogen sulfide ($H_2S$) and carbon dioxide ($CO_2$) volume percent ratio in the acid gas feed stream to sulfur recovery unit. The three facilities are described; the source testing methods used are identified; and the detailed data obtained during the tests program are presented. The emission source tests data that were gathered from other sources for seven additional facilities are also described. The data indicate a range of sulfur recovery efficiencies expected for Claus facilities in relation to the $H_2S$ and $CO_2$ concentrations in the acid gas feed stream and the technology utilized.

## C.1 PROCESS DESCRIPTION OF THE FACILITIES AND THE TESTS RESULTS

### C.1.1 Warren Petroleum's Monument Plant Facility[1]

Warren Petroleum's Monument plant facility is a sour natural gas processing facility that combines removal of associated natural gas liquids, sour natural gas sweetening, and liquid sulfur recovery. The natural gas feed to the plant is produced from the surrounding wells. First, the natural gas liquids are separated, and then acid gas ($H_2S$ and $CO_2$) in the natural gas is separated using an ethanolamine sweetening unit. The acid gas from the sweetening unit is delivered to a Claus sulfur recovery plant. In the Claus process sulfur is recovered from the acid gas via a vapor phase catalytic reaction. The natural gas processed at the facility has a higher concentration of $CO_2$ than $H_2S$, and since both are removed by the sweetening unit, the acid gas feed to the Claus plant is relatively dilute in $H_2S$ content (about 24 percent by volume $H_2S$ in the acid gas feed during the test period). The facility was treating approximately 19.9 $Nm^3/s$ sour natural gas (60.7 MMscf/day) and was producing an average of 18.3 Mg/d liquid sulfur (18 LT/D) during the test period. A simplified flow diagram for the facility is shown in Figure C-1. The Claus plant is a three-stage catalytic unit. Liquid sulfur from the Claus plant is collected in a below-ground storage tank and sold. The Claus plant tail gas is routed to an incinerator, where any remaining $H_2S$ and other reduced sulfur compounds in the tail gas are oxidized to $SO_2$ prior to release into the atmosphere.

Testing of the Claus plant incinerator stack gas was performed to determine the level of $SO_2$, $H_2S$, and total reduced sulfur (TRS) emissions. In addition, the liquid sulfur production rate was monitored in order to determine the sulfur recovery efficiency of the Claus plant. Table C-1 lists the various parameters measured during the testing as well as the sampling and analysis methods used to measure these parameters.

The test results for the $SO_2$, $H_2S$, and TRS concentration levels are summarized in Table C-2. Table C-2 also presents liquid sulfur production rates, $SO_2$ (including TRS expressed as $SO_2$) emission rates and sulfur recovery efficiencies for the Claus plant during the testing period. Figure C-2 graphically presents sulfur recovery efficiency, stack $SO_2$ (including TRS) emission rate, and liquid sulfur production on a daily basis for the testing period. Table C-3 presents information on the daily average velocity, temperature, composition, and actual flow rate of the incinerator stack gas, and Table C-4 presents daily $SO_2$, $H_2S$ (as $SO_2$) and TRS (as $SO_2$) emission rates. The $NO_x$ test results are summarized in Table C-5, and the normal plant operating conditions during the test period are summarized in Table C-6.

Claus plant sulfur recovery efficiency for the facility was calculated based upon the following procedure:

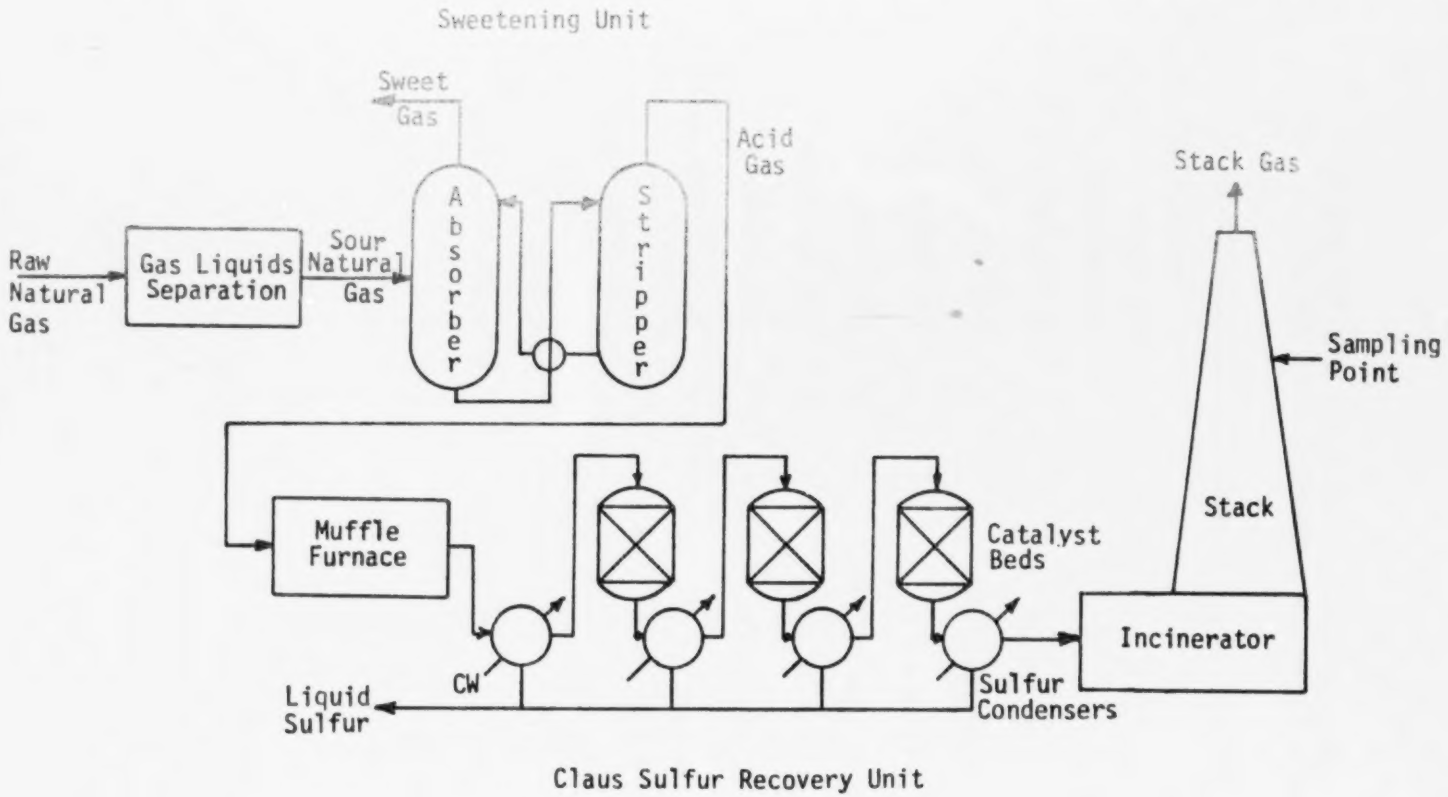
Sweetening Unit
Sweet Gas
Acid Gas
Raw Natural Gas
Gas Liquids Separation
Sour Natural Gas
Absorber
Stripper
Stack Gas
Sampling Point
Stack
Muffle Furnace
Catalyst Beds
Incinerator
CW
Liquid Sulfur
Sulfur Condensers
Claus Sulfur Recovery Unit

Figure C-1. Simplified flow diagram for Warren Petroleum's Monument plant facility.

Table C-1. SAMPLING/ANALYSIS PARAMETERS AND METHODOLOGY AT WARREN PETROLEUM'S MONUMENT PLANT FACILITY

| Measured parameter | Methodology |
|---|---|
| Stack gas volumetric flow rate | EPA Method 2 |
| Stack gas dry molecular weight | EPA Method 3 |
| Stack gas moisture content ($H_2O$) | EPA Method 4 |
| Stack gas sulfur dioxide ($SO_2$) | EPA Method 6 |
| Stack gas nitrogen oxides (as $NO_2$) | EPA Method 7 |
| Stack gas hydrogen sulfide ($H_2S$) | EPA Method 11 |
| Stack gas total reduced sulfur (TRS)[a] | EPA Method 16A |
| Liquid sulfur production | No reference method |

[a] Includes $H_2S$, carbon disulfide ($CS_2$) and carbonyl sulfide (COS).

## Table C-2. WARREN PETROLEUM'S MONUMENT PLANT FACILITY TEST RESULTS SUMMARY

| Test date (1981) | $SO_2$ (ppm) |  | $H_2S$ (ppm) | TRS (ppm) | $SO_2$[a] out stack |  | Sulfur(s)[b] out stack |  | Liquid sulfur production |  | Sulfur recovery efficiency, % |
|---|---|---|---|---|---|---|---|---|---|---|---|
|  | Range | Average |  |  | kg/h | lb/h | Mg/d | LT/D | Mg/d | LT/D |  |
| 3/5 | 3,040-3,580 | 3,300 | c | c | 78.0 | 172.0 | 0.94 | 0.92 | 16.0 | 15.7 | 94.5 |
| 3/6 | 2,900-3,980 | 3,550 | <1 | 21 | 84.0 | 185.1 | 1.01 | 0.99 | 16.8 | 16.5 | 94.3 |
| 3/7 | 3,550-3,730 | 3,640 | <1 | 34 | 89.3 | 196.9 | 1.07 | 1.05 | 18.7 | 18.4 | 94.6 |
| 3/8 | 3,800-4,350 | 4,010 | <1 | 32 | 92.4 | 203.6 | 1.11 | 1.09 | 16.1 | 15.8 | 93.6 |
| 3/9 | 3,390-4,250 | 3,820 | <1 | c | 82.1 | 181.0 | 0.99 | 0.97 | 15.4 | 15.2 | 94.0 |
| 3/10 | 3,190-3,470 | 3,330 | <1 | 22 | 76.3 | 168.1 | 0.91 | 0.90 | 16.4 | 16.1 | 94.7 |
| 3/11 | 3,250-3,420 | 3,340 | <1 | 56 | 74.3 | 163.7 | 0.89 | 0.88 | 16.7 | 16.4 | 94.9 |
| 3/12 | 2,540-3,280 | 2,800 | 1 | 27 | 63.2 | 139.4 | 0.76 | 0.75 | 17.2 | 16.9 | 95.8 |
| 3/13 | 3,730-5,140 | 4,230 | 2 | 24 | 100.3 | 221.2 | 1.21 | 1.19 | 17.5 | 17.2 | 93.5 |
| 3/17 | 3,640-3,740 | 3,690 | <1 | c | 85.3 | 188.0 | 1.03 | 1.01 | 17.8 | 17.5 | 94.6 |
| 3/18 | 3,230-3,590 | 3,440 | <1 | 22 | 84.9 | 187.1 | 1.02 | 1.00 | 19.2 | 18.9 | 95.0 |
| 3/19 | 3,070-3,470 | 3,250 | <1 | 35 | 74.3 | 163.7 | 0.89 | 0.88 | 17.8 | 17.5 | 95.2 |
| 3/20 | 3,390-3,910 | 3,580 | <1 | 26 | 81.8 | 180.3 | 0.99 | 0.97 | 18.4 | 18.1 | 94.9 |
| 3/21 | 3,430-3,580 | 3,500 | <1 | c | 89.4 | 197.0 | 1.08 | 1.06 | 16.6 | 16.3 | 93.9 |
| 3/22 | 3,380-4,430 | 3,770 | <1 | 48 | 89.6 | 197.5 | 1.08 | 1.06 | 18.7 | 18.4 | 94.6 |
| 3/23 | 3,140-3,770 | 3,450 | <1 | 35 | 81.6 | 179.8 | 0.98 | 0.96 | 18.1 | 17.8 | 94.9 |
| 3/24 | 3,440-3,730 | 3,550 | <1 | 25 | 85.9 | 189.3 | 1.03 | 1.01 | 19.3 | 19.0 | 94.9 |
| 3/25 | 3,510-3,560 | 3,540 | <1 | 17 | 84.8 | 186.9 | 1.02 | 1.00 | 19.5 | 19.2 | 95.0 |
| 3/26 | 2,520-2,950 | 2,800 | <1 | 19 | 65.3 | 144.0 | 0.78 | 0.77 | 21.1 | 20.8 | 96.4 |
| 3/27 | 3,150-3,330 | 3,230 | <1 | 15 | 77.5 | 170.8 | 0.94 | 0.92 | 19.5 | 19.2 | 95.4 |
| Test period | 2,520-5,140 | 3,490 | <1 | 29 | 82.0 | 180.8 | 0.99 | 0.97 | 17.8 | 17.5 | 94.8 |

[a] Includes $SO_2$ plus TRS expressed as $SO_2$.

[b] Includes $SO_2$ plus TRS expressed as S.

[c] Test not performed.

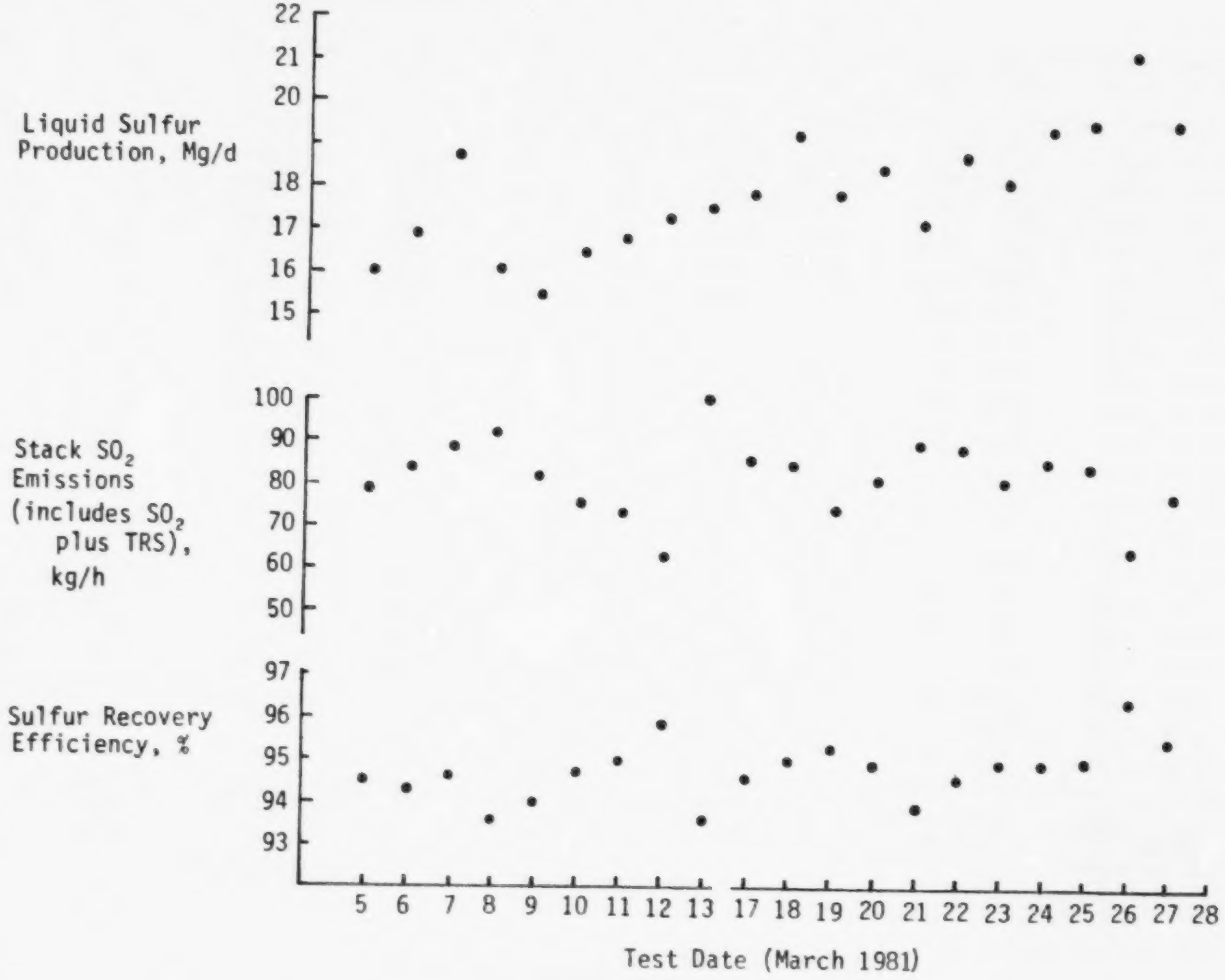

Figure C-2. Summary of Liquid Sulfur Production, Stack $SO_2$ Emissions and Sulfur Recovery Efficiency at Warren Petroleum's Monument Plant Facility.

## Table C-3. WARREN PETROLEUM"S MONUMENT PLANT FACILITY: DAILY AVERAGE STACK GAS VELOCITY, TEMPERATURE, COMPOSITION AND ACTUAL FLOW RATE DURING THE TEST PERIOD

Sampling Location: Incinerator Stack
Stack Diameter (inside): 1.4 m (55 inch)
Pitot Tube Coefficient: 0.84

| Test date (1981) | % $CO_2$ (volume) | % $O_2$ (volume) | % $N_2$ (volume) | % $H_2O$ (volume) | Velocity | | Stack temperature | | Actual flow rate | |
|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | m/s | ft/sec | K | °F | $m^3/s$ | ACFM[a] |
| 3/5 | 24.9 | 8.5 | 66.5 | 15.0 | 6.6 | 21.5 | 901 | 1,163 | 10.0 | 21,200 |
| 3/6 | 25.3 | 7.5 | 65.8 | 15.0 | 6.6 | 21.5 | 899 | 1,158 | 10.0 | 21,200 |
| 3/7 | 25.0 | 8.0 | 66.4 | 15.0 | 6.8 | 22.4 | 899 | 1,158 | 10.4 | 22,100 |
| 3/8 | 25.3 | 8.3 | 66.3 | 15.0 | 6.3 | 20.8 | 899 | 1,158 | 9.7 | 20,600 |
| 3/9 | 26.7 | 7.5 | 65.0 | 15.0 | 5.9 | 19.5 | 896 | 1,154 | 9.1 | 19,300 |
| 3/10 | 26.8 | 7.5 | 65.3 | 15.0 | 6.3 | 20.6 | 899 | 1,159 | 9.6 | 20,400 |
| 3/11 | 25.2 | 8.0 | 66.4 | 15.0 | 6.0 | 19.8 | 894 | 1,150 | 9.3 | 19,600 |
| 3/12 | 25.1 | 8.2 | 66.4 | 15.2 | 6.2 | 20.3 | 894 | 1,150 | 9.5 | 20,100 |
| 3/13 | 26.1 | 7.8 | 66.6 | 15.3 | 6.6 | 21.6 | 894 | 1,150 | 10.1 | 21,300 |
| 3/17 | 26.0 | 7.9 | 66.1 | 15.8 | 6.6 | 21.5 | 900 | 1,160 | 10.1 | 21,300 |
| 3/18 | 25.5 | 8.1 | 66.7 | 13.4 | 6.7 | 22.1 | 894 | 1,150 | 10.3 | 21,800 |
| 3/19 | 26.4 | 8.0 | 66.0 | 14.0 | 6.2 | 20.4 | 904 | 1,168 | 9.5 | 20,200 |
| 3/20 | 25.3 | 8.1 | 66.4 | 14.3 | 6.3 | 20.8 | 898 | 1,156 | 9.7 | 20,600 |
| 3/21 | 25.8 | 8.1 | 66.6 | 13.8 | 7.2 | 23.5 | 899 | 1,159 | 10.9 | 23,200 |
| 3/22 | 25.1 | 8.3 | 66.8 | 14.1 | 6.5 | 21.2 | 898 | 1,156 | 9.9 | 21,000 |
| 3/23 | 26.0 | 8.0 | 66.3 | 13.3 | 6.4 | 21.0 | 900 | 1,160 | 9.8 | 20,700 |
| 3/24 | 26.3 | 7.9 | 66.0 | 14.5 | 6.7 | 21.9 | 899 | 1,158 | 10.2 | 21,600 |
| 3/25 | 26.8 | 7.6 | 65.9 | 15.3 | 6.7 | 22.0 | 900 | 1,160 | 10.2 | 21,700 |
| 3/26 | 27.1 | 7.6 | 65.4 | 15.0 | 6.6 | 21.5 | 898 | 1,157 | 10.0 | 21,200 |
| 3/27 | 27.0 | 7.8 | 65.1 | 15.1 | 6.7 | 22.0 | 897 | 1,155 | 10.2 | 21,700 |

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[a] Calculated at stack gas actual temperature and pressure.

Table C-4. DAILY [a]$SO_2$, $H_2S$ (AS $SO_2$) AND TRS (AS $SO_2$) EMISSIONS DURING THE TEST PERIOD

| Test date (1981) | $SO_2$ emission[a] | | $H_2S$ emission | | TRS emission[b] | |
|---|---|---|---|---|---|---|
| | kg/h | lb/h | kg/h | lb/h | kg/h | lb/h |
| 3/5 | 78.0 | 172.0 | <0.02 | <0.05 | | c |
| 3/6 | 83.0 | 182.9 | <0.02 | <0.05 | 0.5 | 1.1 |
| 3/7 | 87.6 | 193.1 | <0.02 | <0.05 | 0.9 | 1.9 |
| 3/8 | 90.9 | 200.4 | <0.02 | <0.05 | 0.7 | 1.6 |
| 3/9 | 82.1 | 181.0 | <0.02 | <0.05 | | c |
| 3/10 | 75.3 | 165.9 | <0.02 | <0.05 | 0.5 | 1.1 |
| 3/11 | 71.8 | 158.3 | <0.02 | <0.05 | 1.2 | 2.7 |
| 3/12 | 62.0 | 136.6 | <0.03 | <0.06 | 0.6 | 1.4 |
| 3/13 | 99.2 | 218.8 | 0.05 | 0.10 | 0.5 | 1.2 |
| 3/17 | 85.3 | 188.0 | <0.02 | <0.05 | | c |
| 3/18 | 83.9 | 184.9 | <0.02 | <0.05 | 0.5 | 1.1 |
| 3/19 | 72.7 | 160.3 | <0.02 | <0.05 | 0.8 | 1.7 |
| 3/20 | 80.6 | 177.7 | <0.02 | <0.05 | 0.6 | 1.3 |
| 3/21 | 89.4 | 197.0 | <0.03 | <0.06 | | c |
| 3/22 | 87.3 | 192.5 | <0.02 | <0.05 | 1.1 | 2.5 |
| 3/23 | 79.9 | 176.2 | <0.02 | <0.05 | 0.8 | 1.8 |
| 3/24 | 84.7 | 186.7 | <0.02 | <0.05 | 0.6 | 1.3 |
| 3/25 | 84.0 | 185.1 | <0.02 | <0.05 | 0.4 | 0.9 |
| 3/26 | 64.4 | 142.0 | <0.02 | <0.05 | 0.5 | 1.0 |
| 3/27 | 76.7 | 169.2 | <0.02 | <0.05 | 0.4 | 0.8 |

[a] $H_2S$ and TRS not included.

[b] Includes $H_2S$, COS and $CS_2$.

[c] Not available.

## Table C-5. DAILY NO$_x$ TEST RESULTS AND STACK EMISSIONS

| Test Date (1981) | Concentration range (ppm) | Average concentration (ppm) | Average[a] NO$_x$ emissions kg/h | lb/h |
|---|---|---|---|---|
| 3/5 | 6.3 - 11.6 | 8.7 | 0.15 | 0.32 |
| 3/6 | 1.4 - 4.7 | <4.1 | <0.07 | <0.15 |
| 3/7 | 3.0 - 8.5 | 5.3 | 0.10 | 0.21 |
| 3/8 | 10.1 - 10.9 | 10.6 | 0.17 | 0.38 |
| 3/9 | 8.5 - 10.0 | 9.0 | 0.14 | 0.31 |
| 3/10 | 9.4 - 11.2 | 10.7 | 0.17 | 0.38 |
| 3/11 | 10.0 - 11.7 | 10.7 | 0.17 | 0.37 |
| 3/12 | 13.7 - 14.4 | 14.0 | 0.23 | 0.50 |
| 3/13 | 9.4 - 10.3 | 9.7 | 0.16 | 0.36 |
| 3/17 | 11.1 - 13.0 | 12.1 | 0.20 | 0.44 |
| 3/18 | 9.7 - 11.2 | 10.5 | 0.19 | 0.41 |
| 3/19 | 13.2 - 18.1 | 15.3 | 0.25 | 0.55 |
| 3/20 | 11.4 - 13.4 | 12.5 | 0.20 | 0.45 |
| 3/21 | b b | b | b | b |
| 3/22 | 10.7 - 13.9 | 12.3 | 0.21 | 0.46 |
| 3/23 | 13.7 - 14.3 | 14.0 | 0.24 | 0.52 |
| 3/24 | 10.8 - 12.2 | 11.5 | 0.20 | 0.44 |
| 3/25 | 8.6 - 12.5 | 9.9 | 0.17 | 0.38 |
| 3/26 | 9.3 - 10.8 | 12.0 | 0.20 | 0.44 |
| 3/27 | 10.8 - 12.0 | 11.3 | 0.20 | 0.43 |
| Test period | 1.4 - 18.1 | 10.1 | | 11.04 |

[a] NO$_x$ expressed as NO$_2$.

[b] Sample not taken.

## Table C-6. WARREN PETROLEUM'S MONUMENT PLANT FACILITY OPERATING CONDITIONS DURING THE TEST PERIOD

(1) Sour natural gas volumetric flow rate to the sweetening unit:

Average: 19.9 $Nm^3/s$(60.7 MMscf/day)[a]
Maximum: 20.4 $Nm^3/s$(62.2 MMscf/day)
Minimum: 19.4 $Nm^3/s$(59.1 MMscf/day)

(2) Concentration of $H_2S$ in sour natural gas entering the sweetening unit (volume basis):

Average: 0.79%
Maximum: 1.03%
Minimum: 0.43%

(3) Acid gas volumetric flow rate to the Claus unit:

Average: 0.70 $Nm^3/s$(2.12 MMscf/day)
Maximum: 0.75 $Nm^3/s$(2.29 MMscf/day)
Minimum: 0.67 $Nm^3/s$(2.04 MMscf/day)

(4) Acid gas composition (dry volume basis):

| | $H_2S$% | $CO_2$% |
|---|---|---|
| Average: | 24.0 | 76.0 |
| Maximum: | 24.5 | 75.5 |
| Minimum: | 22.5 | 77.5 |

(5) Catalyst bed temperatures (average):

| | Inlet | Outlet |
|---|---|---|
| 1st Reactor | 478 K (400°F) | 587 K (596°F) |
| 2nd Reactor | 478 K (400°F) | 493 K (427°F) |
| 3rd Reactor | 478 K (400°F) | 480 K (405°F) |

(6) Catalyst weights:

| | | |
|---|---|---|
| Reactor Bed #1 | 6,804 kg | (15,000 lbs) |
| Reactor Bed #2 | 8,165 kg | (18,000 lbs) |
| Reactor Bed #3 | 6,804 kg | (15,000 lbs) |

(7) Dates catalyst beds changed:

| | |
|---|---|
| Bed #1 | August 1980 |
| Bed #2 and Bed #3 | March 1977 |

(continued)

Table C-6. Concluded

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(8) Catalyst life expectancy:
3 years to 5 years

(9) Design sulfur recovery efficiency[b]:
94.7%

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[a] Standard conditions: 288.7 K (60°F), $1.01035 \times 10^5$ Pa(29.92 in. Hg).

## (1) FLOW RATES

- Actual cubic meter per second (ACMS)

ACMS = Velocity x Stack Cross Sectional Area

Example: Based on average values of the test dated March 7

$$\text{ACMS} = 6.8\ \frac{\text{meter}}{\text{sec}} \times \frac{(1.4\ \text{m})^2 \times \pi}{4}$$

$\text{ACMS} = 10.4\ \text{m}^3/\text{s}$

- Dry standard cubic meter per second (DSCMS) @ 288.7 K (60°F) and $1.01035 \times 10^5$ Pa (29.92 in. Hg)

$$\text{DSCMS} = \text{ACMS} \times \frac{\text{Barometric Pressure}}{\text{Standard Pressure}} \times \frac{\text{Standard Temp}}{\text{Stack Temp}} \times \begin{matrix}\text{mole}\\ \text{1-fraction}\\ H_2O\end{matrix}$$

Example: Based on average values of the test dated March 7

$$\text{DSCMS} = 10.4\ (\text{ACMS}) \times \frac{0.88473 \times 10^5\ \text{Pa}}{1.01035 \times 10^5\ \text{Pa}} \times \frac{288.7\ \text{K}}{899\ \text{K}} \times (1-0.15)$$

$\text{DSCMS} = 2.5\ \text{m}^3/\text{s}$

## (2) EMISSION RATES

- Emission Rates – $SO_2$ and TRS (kg/h)

$$\text{Emission Rate} = \frac{\text{Concentration of Compound (ppm,dry)}}{10^6} \times \text{DSCMS} \times \frac{\text{compound mole wt}}{\text{molar volume}}$$

Example: Based on average values of the test dated March 7

$$SO_2\ \text{Emission Rate} = \frac{3{,}640\ (\text{ppm}\ SO_2,\text{dry})}{10^6} \times 2.5\ \text{DSCMS} \times \frac{0.064\ \text{kg}\ SO_2}{0.02375\ \text{m}^3} \times \frac{60\ \text{sec}}{\text{min}} \times \frac{60\ \text{min}}{\text{h}}$$

$SO_2$ Emission Rate = 88.3 kg/h

Similarly, for a TRS concentration of 34 ppm dry, the TRS Emission Rate is 0.8 kg/h.

Note: TRS emission rates are expressed as $SO_2$; total $SO_2$ emission rate is therefore the sum of $SO_2$ and TRS emission rates.

### (3) CLAUS PLANT SULFUR RECOVERY EFFICIENCY

$$\text{Recovery Efficiency} = \frac{\text{Sulfur Recovered}}{\text{Sulfur Recovered + Sulfur Emitted}} \times 100\%$$

where:

Sulfur recovered = liquid sulfur production, (Mg/d)
Sulfur emitted = $SO_2$ + TRS emission rates (expressed as elemental sulfur), (Mg/d)

Example: Based on average values of the test dated March 7

Sulfur Recovery Efficiency =

$$\frac{18.7 \text{ Mg/d}}{18.7 \text{ Mg/d} + \left[(88.3 + 0.8)\,\frac{\text{kg } SO_2}{\text{h}} \times \frac{32 \text{ kg S}}{64 \text{ kg } SO_2} \times \frac{\text{Mg}}{1{,}000 \text{ kg}} \times \frac{24 \text{ h}}{\text{d}}\right]} \times 100\%$$

Sulfur Recovery Efficiency = 94.6%

### C.1.2 Getty Oil's New Hope Facility[2]

Getty Oil's New Hope Facility is a sour natural gas processing facility that combines removal of associated natural gas liquids, sour natural gas sweetening, and liquid sulfur recovery. After the natural gas liquids are removed, the natural gas is sweetened using a diglycolamine sweetening unit. Liquid sulfur is recovered from the acid gas generated in the sweetening process by a Claus sulfur recovery unit. The natural gas processed at the facility contains more $H_2S$ than $CO_2$, and therefore, $H_2S$ concentration in the acid gas feed to the Claus plant is relatively high (about 55 percent by volume $H_2S$ in the acid gas feed during the test period). The facility was treating approximately 8.9 $Nm^3/s$ sour natural gas (27.0 MMscf/ day) and was producing an average of 131.1 Mg/d liquid sulfur (128 LT/D) during the test period. A simplified flow diagram for the facility is shown in Figure C-3. The Claus plant is a dual-train, two-stage catalytic unit, with the third catalytic reactor

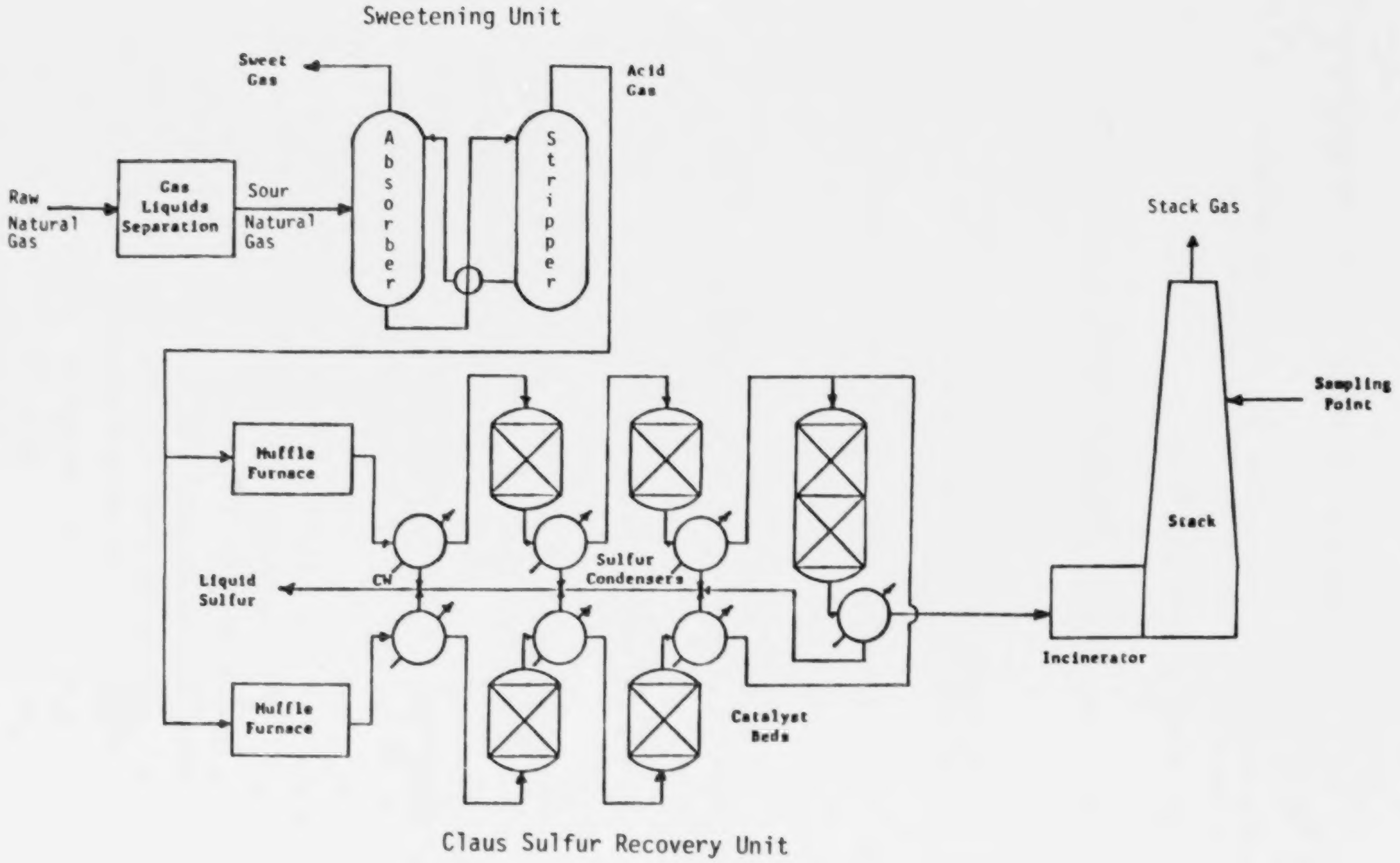
Sweetening Unit
Sweet Gas
Acid Gas
Raw Natural Gas
Gas Liquids Separation
Sour Natural Gas
Absorber
Stripper
Stack Gas
Sampling Point
Stack
Muffle Furnace
Sulfur Condensers
Liquid Sulfur
CW
Incinerator
Muffle Furnace
Catalyst Beds
Claus Sulfur Recovery Unit

Figure C-3. Simplified flow diagram for Getty Oil's New Hope Facility.

being common to both the trains. Liquid sulfur from the Claus plant is collected in a below-ground storage tank and sold. The Claus plant tail gas is routed to an incinerator to oxidize the residual $H_2S$ and other reduced sulfur compounds to $SO_2$ prior to emission to the atmosphere.

Testing of the Claus plant incinerator stack gas was performed to determine the level of $SO_2$, $H_2S$, and TRS emissions from the stack. In addition, the liquid sulfur production rate was monitored in order to determine the sulfur recovery efficiency of the Claus plant. The various parameters measured during the testing at the facility and the sampling and analysis methods used to measure these parameters are listed in Table C-7.

The test results for the $SO_2$, $H_2S$, and TRS concentration levels are summarized in Table C-8. Table C-8 also presents liquid sulfur production rates, $SO_2$ (including TRS expressed as $SO_2$) emission rates and sulfur recovery efficiencies for the Claus plant during the testing period. Figure C-4 graphically presents the sulfur recovery efficiency, stack $SO_2$ (including TRS) emission rate, and liquid sulfur production on a daily basis for the testing period. Table C-9 presents information on the daily average velocity, temperature, composition, and actual flow rate of the incinerator stack gas and Table C-10 presents daily $SO_2$, $H_2S$ (as $SO_2$) and TRS (as $SO_2$) emission rates. The $NO_x$ test results are summarized in Table C-11, and the normal plant operating conditions during the test period are listed in Table C-12.

Claus plant sulfur recovery efficiency for the facility was calculate based upon the following procedure:

(1) FLOW RATES

o Actual cubic meter per second (ACMS)

ACMS = Velocity x Stack Cross Sectional Area

Example: Based on average values of the test dated April 8

$$\text{ACMS} = 17.2\ \frac{\text{meter}}{\text{sec}} \times \frac{(1.2446\ \text{m})^2 \times \pi}{4}$$

Table C-7. SAMPLING/ANALYSIS PARAMETERS AND METHODOLOGY AT GELL OIL'S NEW HOPE FACILITY

| Measured parameter | Methodology |
|---|---|
| Stack gas volumetric flow rate | EPA Method 2 |
| Stack gas dry molecular weight | EPA Method 3 |
| Stack gas moisture content ($H_2O$) | EPA Method 4 |
| Stack gas sulfur dioxide ($SO_2$) | EPA Method 6 |
| Stack gas nitrogen oxides (as $NO_2$) | EPA Method 7 |
| Stack gas hydrogen sulfide ($H_2S$) | EPA Method 11 |
| Stack gas total reduced sulfur (TRS)[a] | EPA Method 16A |
| Liquid sulfur production | No reference method |

[a] Includes $H_2S$, $CS_2$ and COS.

## Table C-8. GETTY OIL'S NEW HOPE FACILITY TEST RESULTS SUMMARY

| Test date (1981) | $SO_2$ (ppm) | | $H_2S$ (ppm) | | TRS (ppm) | Stack $SO_2$ [a] | | Stack Sulfur (S) [b] | | Liquid sulfur production | | Sulfur recovery efficiency, % |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | Range | Average | Range | Average | | kg/h | lb/h | Mg/d | LT/D | Mg/d | LT/D | |
| 4/8 | 8,300 - 9,840 | 8,950 | 425 - 662 | 545 | 1,050 | 593 | 1,318 | 7.0 | 7.1 | 144.7 | 147.0 | 95.4 |
| 4/9 | 8,950 - 13,600 | 10,520 | 446 - 925 | 637 | 1,460 | 710 | 1,565 | 8.4 | 8.5 | 146.2 | 148.5 | 94.6 |
| 4/10 | 9,360 - 10,300 | 9,950 | 333 - 1,800 | 959 | 977 | 536 | 1,181 | 6.3 | 6.4 | 123.2 | 125.2 | 95.1 |
| 4/11 | 8,930 - 9,370 | 9,080 | 331 - 402 | 359 | 657 | 506 | 1,116 | 6.0 | 6.1 | 113.0 | 114.8 | 95.0 |
| 4/12 | 8,850 - 9,480 | 9,150 | 144 - 408 | 244 | 787 | 518 | 1,141 | 6.1 | 6.2 | 113.0 | 114.8 | 94.9 |
| Test period | 8,300 - 13,600 | 9,517 | 144 - 1,800 | 549 | 972 | 573 | 1,264 | 6.8 | 6.9 | 128.0 | 130.1 | 95.0 |

[a] Includes $SO_2$ plus TRS expressed as $SO_2$.

[b] Includes $SO_2$ plus TRS expressed as S.

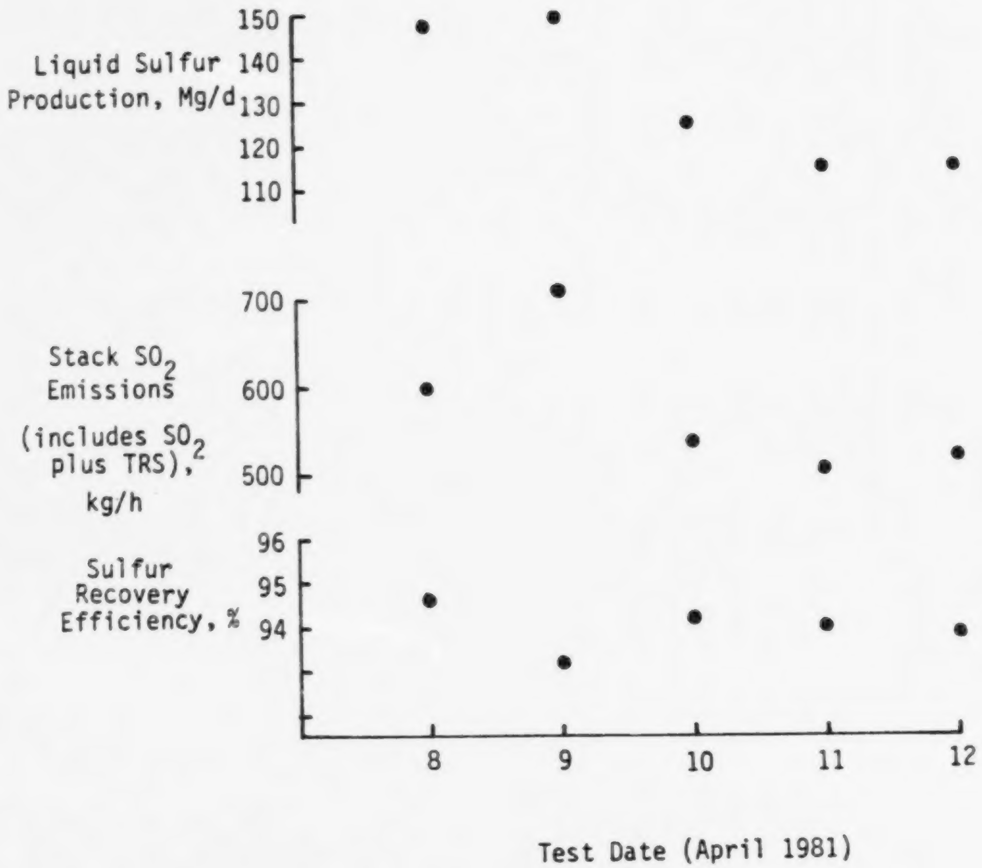

Figure C-4. Summary of Liquid Sulfur Production, Stack $SO_2$ Emissions and Sulfur Recovery Efficiency at Getty Oil's New Hope Facility.

## Table C-9. GETTY OIL'S NEW HOPE FACILITY: DAILY AVERAGE STACK GAS VELOCITY, TEMPERATURE, COMPOSITION AND ACTUAL FLOW RATE DURING THE TEST PERIOD

Sampling Location: Incinerator Stack
Stack Diameter (inside): 1.24 m (49 inches)
Pitot Tube Coefficient: 0.84

| Test date (1981) | % $CO_2$ (volume) | % $O_2$ (volume) | % $N_2$ (volume) | % $H_2O$ (volume) | Velocity | | Stack temperature | | Actual flow rate | |
|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | m/s | ft/sec | K | °F | $m^3/s$ | ACFM[a] |
| 4/8 | 16.2 | 5.4 | 75.3 | 26.6 | 17.2 | 56.5 | 725 | 845 | 20.9 | 44,300 |
| 4/9 | 16.9 | 4.8 | 75.6 | 27.5 | 17.6 | 57.6 | 729 | 852 | 21.3 | 45,200 |
| 4/10 | 19.2 | 3.0 | 74.3 | 29.6 | 14.1 | 46.4 | 694 | 790 | 17.2 | 36,400 |
| 4/11 | 17.3 | 5.3 | 74.3 | 26.3 | 15.5 | 51.0 | 753 | 895 | 18.9 | 40,000 |
| 4/12 | 17.1 | 5.4 | 74.3 | 26.9 | 15.6 | 51.3 | 750 | 890 | 19.0 | 40,300 |

[a] Calculated at stack gas actual temperature and pressure.

## Table C-10. DAILY [a]$SO_2$, $H_2S$ (AS $SO_2$) AND TRS (AS $SO_2$) EMISSIONS DURING THE TEST PERIOD

| Test date (1981) | $SO_2$ emission[a] |  | $H_2S$ emission |  | TRS emission[b] |  |
|---|---|---|---|---|---|---|
|  | kg/h | lb/h | kg/h | lb/h | kg/h | lb/h |
| 4/8 | 535 | 1,180 | 33 | 72 | 63 | 138 |
| 4/9 | 621 | 1,370 | 39 | 87 | 88 | 195 |
| 4/10 | 490 | 1,080 | 46 | 101 | 46 | 101 |
| 4/11 | 472 | 1,040 | 19 | 41 | 34 | 76 |
| 4/12 | 476 | 1,050 | 13 | 28 | 41 | 91 |

[a] $H_2S$ and TRS not included.

[b] Includes $H_2S$, COS and $CS_2$.

Table C-11. DAILY $NO_x$ TEST RESULTS AND STACK EMISSIONS

| Test date (1981) | Concentration range (ppm) | Average concentration (ppm) | Average $NO_x$[a] emissions | |
|---|---|---|---|---|
| | | | kg/h | lb/h |
| 4/8 | <3.0 - 13.0 | <6.3 | <0.3 | <0.6 |
| 4/9 | <3.0 - <13.0 | <6.3 | <0.3 | <0.6 |
| 4/10 | <3.0 | <3.0 | <0.1 | <0.3 |
| 4/11 | <3.0 | <3.0 | <0.1 | <0.3 |
| 4/12 | <3.0 | <3.0 | <0.1 | <0.3 |

[a] $NO_x$ expressed as $NO_2$.

## Table C-12. GETTY OIL'S NEW HOPE FACILITY OPERATING CONDITIONS DURING THE TEST PERIOD

(1) Sour natural gas volumetric flow rate to the sweetening unit:

Average: 8.9 $Nm^3/s$ (27.0 MMscf/day)[a]
Maximum: 9.2 $Nm^3/s$ (28.0 MMscf/day)
Minimum: 8.6 $Nm^3/s$ (26.1 MMscf/day)

(2) Concentration of $H_2S$ in sour natural gas entering the sweetening unit (volume basis):

Average: 7%
Range: 6.5% - 10%

(3) Acid gas volumetric flow rate to the Claus unit:

Average: 2.0 $Nm^3/s$ (6.1 MMscf/day)
Maximum: 2.1 $Nm^3/s$ (6.4 MMscf/day)
Minimum: 1.9 $Nm^3/s$ (5.7 MMscf/day)

(4) Acid gas composition (dry basis):

| | $H_2S$% | $CO_2$% |
|---|---|---|
| Average: | 55.0 | 45.0 |
| Maximum: | 55.6 | 44.4 |
| Minimum: | 54.4 | 45.6 |

(5) Catalyst bed temperatures:

| | Inlet | Outlet |
|---|---|---|
| 1st Train: | | |
| 1st Reactor | 490 K (423°F) | 555 K (539°F) |
| 2nd Reactor | 504 K (447°F) | 518 K (473°F) |
| 2nd Train: | | |
| 1st Reactor | 490 K (422°F) | 565 K (557°F) |
| 2nd Reactor | 505 K (450°F) | 530 K (495°F) |
| Common Bed Reactor | Average of Inlet and Outlet: | |
| 3rd Reactor | 500 K (400°F) | |

(6) Dates catalyst beds changed:

| | |
|---|---|
| 1st Train (1st Reactor and 2nd Reactor) | August 1980 |
| 2nd Train (1st Reactor and 2nd Reactor) | August 1980 |
| Common Bed Reactor | 1972 |

(7) Design sulfur recovery efficiency[b]:

96%

---

[a] Standard conditions: 288.7 K (60°F), 1.01035 x $10^5$ Pa (29.92 in. Hg).

[b] Supplied by the facility.

ACMS = 20.9 $m^3/s$

- Dry standard cubic meter per second (DSCMS) @ 288.7 K (60°F) and $1.01035 \times 10^5$ Pa (29.92 in. Hg)

$$\text{DSCMS} = \text{ACMS} \times \frac{\text{Barometric Pressure}}{\text{Standard Pressure}} \times \frac{\text{Standard Temp}}{\text{Stack Temp}} \times \begin{matrix}\text{mole}\\ \text{1-fraction}\\ H_2O\end{matrix}$$

Example: Based on average values of the test dated April 8

$$\text{DSCMS} = 20.9\ (\text{ACMS}) \times \frac{1.011363 \times 10^5\ \text{Pa}}{1.01035 \times 10^5\ \text{Pa}} \times \frac{288.7\ \text{K}}{725\ \text{K}} \times (1-0.2660)$$

DSCMS = 6.1 $m^3/s$

(2) EMISSION RATES

- Emission rates – $SO_2$ and TRS (kg/h)

$$\text{Emission Rate} = \frac{\text{Concentration of Compound (ppm,dry)}}{10^6} \times \text{DSCMS} \times \frac{\text{Compound Mole wt}}{\text{molar volume}}$$

Example: Based on average values of the test dated April 8

$$SO_2\ \text{Emission Rate} = \frac{8{,}950\ (\text{ppm}\ SO_2,\ \text{dry})}{10^6} \times 6.1\ m^3/S \times \frac{0.064\ \text{kg}\ SO_2}{0.02375\ m^3} \times \frac{60\ \text{sec}}{\text{min}} \times \frac{60\ \text{min}}{\text{h}}$$

$SO_2$ Emission Rate = 529.6 kg/h

Similarly for a TRS concentration of 1,050 ppm dry, the TRS emission rate is 62.3 kg/h.

Note: TRS emission rates are expressed as $SO_2$; total $SO_2$ emission rate is therefore the sum of $SO_2$ and TRS emission rates.

(3) CLAUS PLANT SULFUR RECOVERY EFFICIENCY

$$\text{Recovery Efficiency} = \frac{\text{Sulfur Recovered}}{\text{Sulfur Recovered} + \text{Sulfur Emitted}} \times 100\%$$

where:

Sulfur recovered = liquid sulfur production, Mg/d
Sulfur emitted = $SO_2$ + TRS emission rates (both expressed as elemental sulfur), Mg/d

Example: Based on average values of the test dated April 8

Sulfur Recovery Efficiency =

$$\frac{147.0\ \text{Mg/d}}{147.0\ \text{Mg/d} + \left[(529.6\ \text{kg/h}\ SO_2 + 62.3\ \text{kg/h TRS}) \times \frac{32\ \text{kg S}}{64\ \text{kg}\ SO_2} \times \frac{\text{Mg}}{1{,}000\ \text{kg}} \times \frac{24\ \text{h}}{\text{d}}\right]} \times 100\%$$

Sulfur Recovery Efficiency = 95.4%

### C.1.3 Shell Oil's Thomasville Facility[3]

Shell Oil's Thomasville facility is a sour natural gas processing facility engaged in sweetening of the sour gas and recovering sulfur from the acid gas generated in the sweetening operation. As the gas is dry, no liquids are associated with the gas. Natural gas feed to the plant is produced from six local gas wells. The gas is sweetened in a Sulfinol sweetening unit. The Sulfinol unit was treating about 28.5 $Nm^3/s$ sour natural gas (87 MMscf/day) during the test period. The acid gas stream from the Sulfinol unit is relatively rich in $H_2S$ (about 84.4 percent by volume $H_2S$ in the acid gas feed during the test period). The acid gas stream is fed to a Claus sulfur recovery plant. The Claus plant was producing an average of 1,174 Mg/d liquid sulfur (1,155 LT/D) during the test period. A simplified flow diagram of the facility is shown in Figure C-5. The Claus plant is a three-stage catalytic unit. A portion of the acid gas feed stream bypasses the reaction furnace and is used to fuel the three in-line burners (for the purpose of reheating). Liquid sulfur from the Claus plant is collected in a below-ground storage tank and sold. The Claus plant tail gas is routed to an incinerator to oxidize any residual $H_2S$ and other reduced sulfur compounds to $SO_2$ prior to release to the atmosphere.

Testing of the Claus plant incinerator stack gas was performed to determine the level of $SO_2$, $H_2S$, and TRS emissions. In addition, the liquid sulfur production rate was monitored in order to determine sulfur

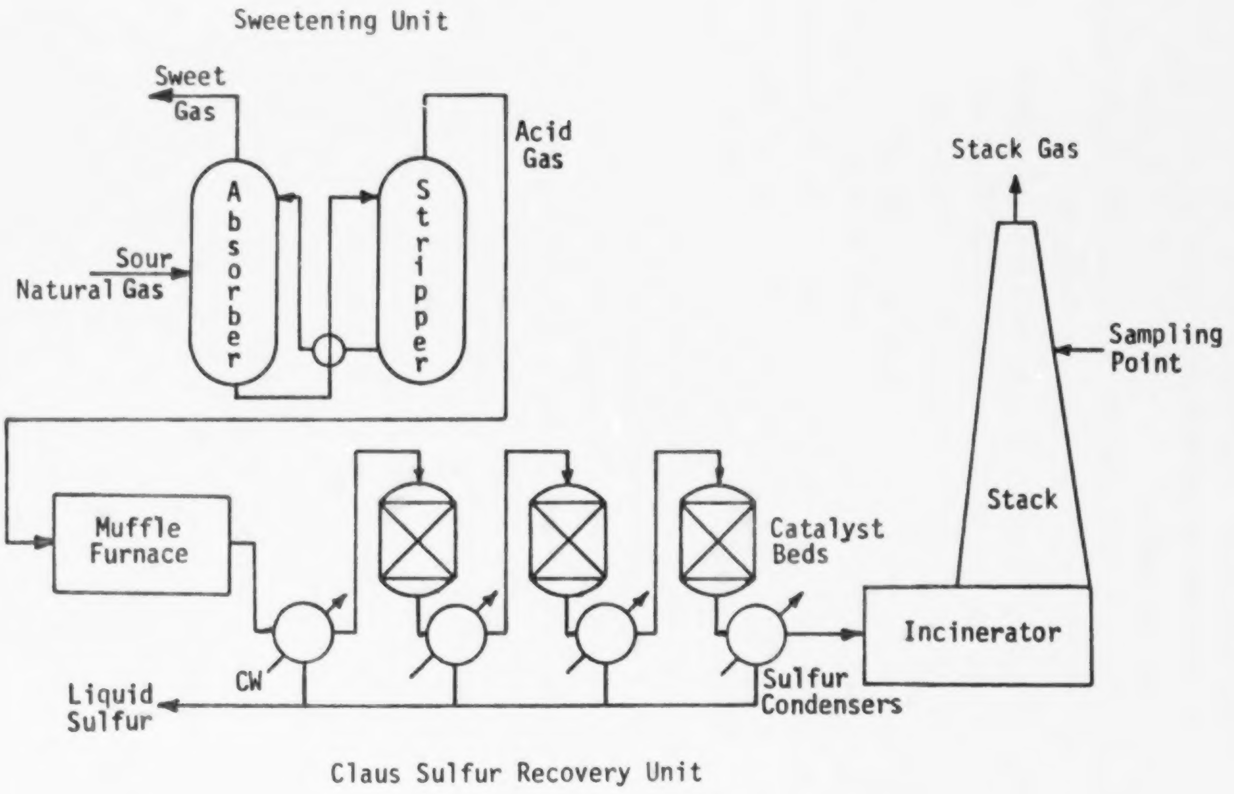
Sweetening Unit
Sweet Gas
Acid Gas
Sour Natural Gas
Absorber
Stripper
Stack Gas
Sampling Point
Stack
Muffle Furnace
Catalyst Beds
Incinerator
CW
Liquid Sulfur
Sulfur Condensers
Claus Sulfur Recovery Unit

Figure C-5. Simplified flow diagram for Shell Oil's Thomasville Facility.

recovery efficiency of the Claus plant. Table C-13 lists the various parameters measured during the testing and the sampling and analysis methods used to measure these parameters.

The test results for the $SO_2$, $H_2S$, and TRS concentration levels are summarized in Table C-14. Table C-14 also presents liquid sulfur production rates, $SO_2$ (including TRS expressed as $SO_2$) emission rates, and sulfur recovery efficiencies for the Claus plant during the test period. Figure C-6 graphically presents sulfur recovery efficiency, stack $SO_2$ (including TRS) emission rate, and liquid sulfur production on a daily basis for the test period. Table C-15 presents the daily average velocity, temperature, composition and actual flow rate of the incinerator stack gas and Table C-16 presents daily $SO_2$, $H_2S$ (as $SO_2$) and TRS (as $SO_2$) emission rates. The $NO_x$ test results are summarized in Table C-17. Table C-18 lists normal plant operating conditions during the test period. In addition, Table C-19 shows a comparison of the test results obtained by the EPA/Emissions Measurement Branch and the results of the tests conducted at the same time by the company's operations support group trailer laboratory.

Claus plant sulfur recover efficiency for the facility was calculated based upon the following procedure:

(1) FLOW RATES

- Actual cubic meter per second (ACMS)

ACMS = Velocity x Stack Cross Sectional Area

Example: Based on average values of the test dated May 12

$$\text{ACMS} = 23.0\ \frac{\text{meter}}{\text{sec}} \times \frac{(2.896\text{m})^2 \times \pi}{4}$$

$$\text{ACMS} = 151.5\ \text{m}^3/\text{s}$$

- Dry standard cubic meter per second (DSCMS) @ 288.7 K (60°F) and 1.01035 x $10^5$ Pa (29.92 in. Hg)

$$\text{DSCMS} = \text{ACMS} \times \frac{\text{Barometric Pressure}}{\text{Standard Pressure}} \times \frac{\text{Standard Temp}}{\text{Stack Temp}} \times \begin{matrix}\text{mole}\\ \text{1-fraction}\\ H_2O\end{matrix}$$

## Table C-13. SAMPLING/ANALYSIS PARAMETERS AND METHODOLOGY AT SHELL OIL'S THOMASVILLE FACILITY

| Measured parameter | Methodology |
|---|---|
| Stack gas volumetric flow rate | EPA Method 2 |
| Stack gas dry molecular weight | EPA Method 3 |
| Stack gas moisture content ($H_2O$) | EPA Method 4 |
| Stack gas sulfur dioxide ($SO_2$) | EPA Method 6 |
| Stack gas nitrogen oxides (as $NO_2$) | EPA Method 7 |
| Stack gas hydrogen sulfide ($H_2S$) | EPA Method 11 |
| Stack gas total reduced sulfur (TRS)[a] | EPA Method 16A |
| Liquid sulfur production | No reference method |

[a] Includes $H_2S$, $CS_2$ and COS.

Table C-14. SHELL OIL'S THOMASVILLE FACILITY TEST RESULTS SUMMARY

| Test date (1981) | $SO_2$ (ppm) |  | $H_2S$ (ppm) |  | TRS (ppm) | Stack $SO_2$[a] |  | Stack sulfur[b] |  | Liquid sulfur production |  | Sulfur recovery efficiency, % |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
|  | range | average | range | average |  | kg/h | lb/h | Mg/d | LT/D | Mg/d | LT/D |  |
| 5/12 | 8,764- 9,161 | 8,907 | 13-30 | 19 | 186 | 3,168 | 6,984 | 38.0 | 37.4 | 1,241 | 1,221 | 97.0 |
| 5/13 | 8,792- 9,352 | 9,136 | 20-47 | 36 | 314 | 3,350 | 7,386 | 40.2 | 39.6 | 1,116 | 1,098 | 96.5 |
| 5/14 | 8,615- 8,990 | 8,779 | 43-48 | 46 | 330 | 3,162 | 6,972 | 38.0 | 37.4 | 1,192 | 1,173 | 96.9 |
| 5/15 | 8,522- 8,704 | 8,616 | 16-23 | 20 | 245 | 3,078 | 6,786 | 37.0 | 36.4 | 1,189 | 1,170 | 97.0 |
| 5/18 | 9,091-10,125 | 9,671 | 23-33 | 28 | 291 | 3,475 | 7,662 | 41.7 | 41.0 | 1,129 | 1,111 | 96.4 |
| Test Period | 8,522-10,125 | 9,022 | 13-48 | 30 | 273 | 3,239 | 7,140 | 39.0 | 38.4 | 1,174 | 1,155 | 96.8 |

[a] Includes $SO_2$ plus TRS expressed as $SO_2$.

[b] Includes $SO_2$ plus TRS expressed as S.

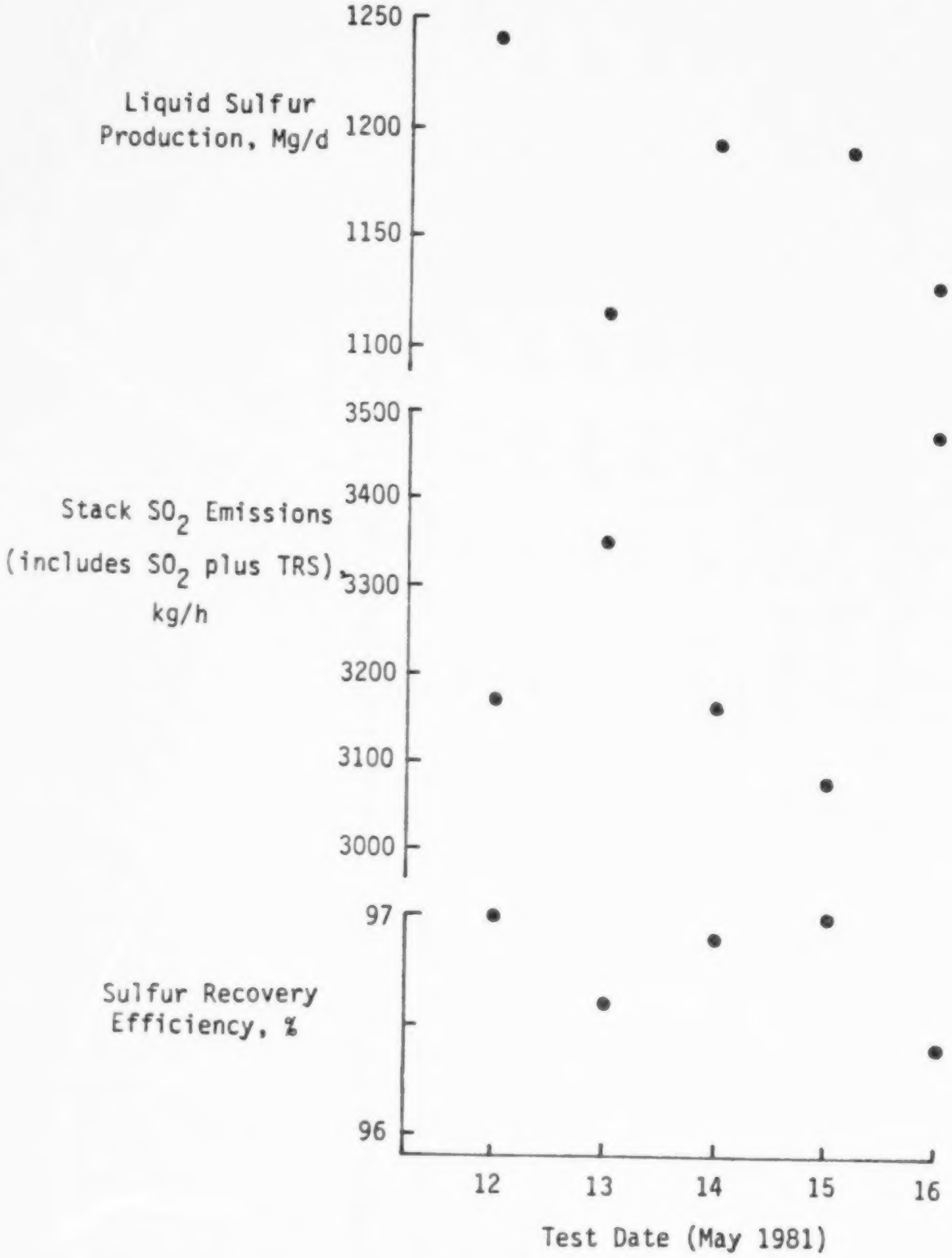

Figure C-6. Summary of Liquid Sulfur Production, Stack $SO_2$ Emissions, and Sulfur Recovery Efficiency at Shell Oil's Thomasville Facility.

Table C-15. SHELL OIL'S THOMASVILLE FACILITY: DAILY AVERAGE STACK GAS VELOCITY, TEMPERATURE, COMPOSITION AND ACTUAL FLOW RATE DURING THE TEST PERIOD

Sampling Location: Incinerator Stack
Stack Diameter (inside): 2.896 m (9.5 ft)
Pitot Tube Coefficient: 0.84

| Test date (1981) | % $CO_2$ (volume) | % $O_2$ (volume) | % $N_2$ (volume) | % $H_2O$ (volume) | Velocity | | Stack temperature | | Actual flow rate | |
|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | m/s | ft/sec | K | °F | $m^3/s$ | ACFM[a] |
| 5/12 | 8.1 | 2.6 | 88.6 | 28.39 | 23.0 | 75.5 | 876 | 1,118 | 151.5 | 321,000 |
| 5/13 | 7.9 | 3.2 | 88.4 | 27.81 | 22.7 | 74.4 | 845 | 1,062 | 148.7 | 315,000 |
| 5/14 | 8.4 | 2.5 | 88.5 | 28.65 | 23.3 | 76.4 | 876 | 1,117 | 153.4 | 325,000 |
| 5/15 | 8.3 | 2.5 | 88.6 | 27.94 | 23.3 | 76.5 | 887 | 1,137 | 153.9 | 326,000 |
| 5/18 | 8.6 | 2.0 | 88.7 | 29.22 | 23.7 | 77.8 | 888 | 1,138 | 156.2 | 331,000 |

[a] Calculated at stack gas actual temperature and pressure.

## Table C-16. DAILY $SO_2$, $H_2S$ (AS $SO_2$) AND TRS (AS $SO_2$) EMISSIONS DURING THE TEST PERIOD

| Test date (1981) | $SO_2$ emission[a] kg/h | $SO_2$ emission[a] lb/h | $H_2S$ emission kg/h | $H_2S$ emission lb/h | TRS emission[b] kg/h | TRS emission[b] lb/h |
|---|---|---|---|---|---|---|
| 5/12 | 3,103 | 6,840 | 6.5 | 14.4 | 65.3 | 144 |
| 5/13 | 3,239 | 7,140 | 12.8 | 28.2 | 111.6 | 246 |
| 5/14 | 3,048 | 6,720 | 16.1 | 35.4 | 114.3 | 252 |
| 5/15 | 2,985 | 6,580 | 7.1 | 15.6 | 84.4 | 206 |
| 5/18 | 3,375 | 7,440 | 9.8 | 21.6 | 100.7 | 222 |

[a] $H_2S$ and TRS not included.

[b] Includes $H_2S$, COS and $CS_2$.

Table C-17. DAILY $NO_x$ TEST RESULTS AND STACK EMISSIONS

| Testing date (1981) | Concentration range (ppm) | Average concentration (ppm) | Average $NO_x$[a] emissions kg/h | lb/h |
|---|---|---|---|---|
| 5/12 | <3.0-6.2 | <4.2 | <1.0 | <2.3 |
| 5/13 | <3.0 | <3.0 | <0.8 | <1.7 |
| 5/14 | <3.0-3.7 | <3.2 | <0.8 | <1.7 |
| 5/15 | <3.0-4.7 | <3.7 | <0.9 | <2.0 |
| 5/18 | <3.0-5.4 | <3.8 | <1.0 | <2.1 |

[a] $NO_x$ expressed as $NO_2$.

## Table C-18. SHELL OIL'S THOMASVILLE FACILITY OPERATING CONDITIONS DURING THE TEST PERIOD

(1) Sour natural gas volumetric flow rate to the sweetening unit:

Average: 28.5 $Nm^3/s$(87.0 MMscf/day)[a]
Maximum: 28.7 $Nm^3/s$(87.7 MMscf/day)
Minimum: 28.5 $Nm^3/s$(86.8 MMscf/day)

(2) Concentration of $H_2S$ in sour natural gas (combined from all the six wells) entering the sweetening unit: (% $H_2S$ from the individual gas wells ranges from 28.8% to 45.3%)

Average: 35.1%

(3) Acid gas volumetric flow rate to the Claus unit:

Average: 11.9 $Nm^3/s$(36.4 MMscf/day)
Maximum: 12.2 $Nm^3/s$(37.1 MMscf/day)
Minimum: 11.8 $Nm^3/s$(36.1 MMscf/day)

(4) Acid gas composition (dry basis):

| | $H_2S$% | $CO_2$% |
|---|---|---|
| Average: | 84.4 | 15.6 |
| Maximum: | 85.4 | 14.6 |
| Minimum: | 84.0 | 16.0 |

(5) Catalyst bed temperatures (average):

| | Inlet | Outlet |
|---|---|---|
| 1st Reactor | 497 K (435°F) | 603 K (625°F) |
| 2nd Reactor | 481 K (407°F) | 510 K (459°F) |
| 3rd Reactor | 479 K (403°F) | 486 K (415°F) |

(6) Catalyst (Kaiser S-201) weights; dimensions:

| | | |
|---|---|---|
| Reactor Bed #1 | 102,285 kg | (225,500 lbs) |
| Reactor Bed #2 | 102,285 kg | (225,500 lbs) |
| Reactor Bed #3 | 102,285 kg | (225,500 lbs) |
| Each Bed Volume | 138.8 $m^3$ | (4,902 cubic feet) |
| Each Bed Diameter | 12.0 m | (39.5 feet) |
| Each Bed Thickness | 1.22 m | (4 feet) |

(7) Dates catalyst beds changed:

Bed #1, #2 and #3 April, 1979

(continued)

Table C-18. Concluded

(8) Individual bed efficiency (approximate):

| Bed | Efficiency[b] |
|---|---|
| #1 | 72% |
| #2 | 68% |
| #3 | 19% |

(9) Catalyst normal life expectancy:

5 years (3 years - 5 years range)

(10) Design sulfur recovery efficiency[b]: 97.83%

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[a] Standard conditions: 288.7 K (60°F), 1.01035 x $10^5$ Pa (29.92 in. Hg)

[b] Supplied by the facility.

Table C-19. COMPARISON OF EPA EMISSIONS MEASUREMENT BRANCH (EMB) TESTS RESULTS AND THE COMPANY'S OPERATIONS SUPPORT LABORATORY TEST RESULTS AT THE THOMASVILLE FACILITY

|  | Test date (1981) | % $CO_2$ (volume) | % $O_2$ (volume) | % $N_2$ (volume) | $SO_2$ (ppm) | $H_2S$ (ppm) | TRS (ppm) | Liquid sulfur production Mg/d | Liquid sulfur production LT/d | Sulfur recovery efficiency (%) |
|---|---|---|---|---|---|---|---|---|---|---|
| EPA/EMB's Tests | 5/12 | 8.2 | 2.5 | 88.6 | 9,161 | 30 | 186 | 1,241 | 1,221 | 97.03 |
| Tests by Company's Operations Support Lab | 5/12 | 7.7 | 3.5 | 86.4 | 9,460 | -- | 80 | 1,164 | 1,146 | 96.99 |
| EPA/EMB's Tests | 5/13 | 7.4 | 3.9 | 88.4 | 9,352 | 40 | 314 | 1,116 | 1,098 | 96.52 |
| Tests by Company's Operations Support Lab | 5/13 | 7.8 | 3.3 | 86.3 | 9,620 | 56 | 166 | 1,178 | 1,159 | 97.23 |

Example: based on average values of the test dated May 12

$$\text{DSCMS} = 151.5\ (\text{ACMS}) \times \frac{1.011363 \times 10^5\ \text{Pa}}{1.01035 \times 10^5\ \text{Pa}} \times \frac{288.7\ \text{K}}{876\ \text{K}} \times (1-0.2839)$$

$$\text{DSCMS} = 35.8\ \text{m}^3/\text{s}$$

(2) EMISSION RATES

- Emission Rates – $SO_2$ and TRS (kg/h)

$$\text{Emission Rate} = \frac{\text{Concentration of Compound (ppm,dry)}}{10^6} \times \text{DSCMS} \times \frac{\text{compound mole wt}}{\text{molar volume}}$$

Example: Based on average values of the test dated May 12

$$SO_2\ \text{Emission Rate} = \frac{8907\ (\text{ppm } SO_2,\text{dry})}{10^6} \times 35.8\ \text{m}^3/\text{s} \times \frac{0.064\ \text{kg } SO_2}{0.02375\ \text{m}^3} \times \frac{60\ \text{sec}}{\text{min}} \times \frac{60\ \text{min}}{\text{h}}$$

$$SO_2\ \text{Emission Rate} = 3{,}093.4\ \text{kg/h}$$

Similarly, for a TRS concentration of 186 ppm dry, the TRS Emission Rate is 64.8 kg/h

Note: TRS emission rates are expressed as $SO_2$; total $SO_2$ emission rate is therefore the sum of $SO_2$ and TRS emission rates.

(3) CLAUS PLANT SULFUR RECOVERY EFFICIENCY

$$\text{Recovery Efficiency} = \frac{\text{Sulfur Recovered}}{\text{Sulfur Recovered} + \text{Sulfur Emitted}} \times 100\%$$

where:

Sulfur recovered = liquid sulfur production, (Mg/d)
Sulfur emitted = $SO_2$ + TRS emission rates (expressed as elemental sulfur), (Mg/d)

Example: Based on average values of the test dated May 12

Sulfur Recovery Efficiency =

$$\frac{1{,}241\ \text{Mg/d}}{1{,}241\ \text{Mg/d} + [(3{,}093.4\ \text{kg/h } SO_2 + 64.8\text{kg/h TRS}) \times \frac{32\ \text{kg S}}{64\ \text{kg } SO_2} \times \frac{\text{Mg}}{1{,}000\ \text{kg}} \times \frac{24\ \text{h}}{\text{d}}]} \times 100\%$$

Sulfur Recovery Efficiency = 97.0%

### C.1.4 Exxon's Blackjack Creek Facility[4]

The Blackjack Creek facility is a sour natural gas processing facility that employs a three-stage Claus sulfur recovery plant with a SCOT tail gas cleanup unit. A simplified flow diagram for the facility is shown in Figure C-7. Acid gas from sweetening unit flows to the Claus plant at an average rate of 1.134 $Nm^3/s$ (3.459 MMscf/day) with an average hydrogen sulfide concentration of 85.8 percent by volume. The facility recovers approximately 105.2 Mg/d liquid sulfur (103.6 LT/D) and normally operates at 99.8 percent recovery efficiency. Residual tail gas from the SCOT unit is routed to an incinerator where the remaining $H_2S$ and other reduced sulfur compounds are oxidized to $SO_2$ prior to emission to the atmosphere. Table C-20 presents annual stack emissions and sulfur recovery efficiency tests data from 1977 through 1980 that were supplied by the facility. Sulfur recovery efficiency is calculated as sulfur inlet minus stack sulfur emission, divided by sulfur inlet.

### C.1.5 Shell Oil' Bryans Mill Facility[5]

The Bryans Mill facility is a Claus sulfur recovery facility that began operation as a two-stage plant in 1962 and was expanded to a three-stage plant in September 1967. During 1975, the average inlet sour natural gas flow rate was 21.2 $Nm^3/s$ (64.7 MMscf/day), the average liquid sulfur production was 191 Mg/d (188 LT/D) and $H_2S$ concentration in the sour natural gas averaged 7.8 percent by volume. A summary of the source emission tests conducted during 1973 and 1974 is presented in Table C-21. During the period from March 1972 through June 1975, the $H_2S/CO_2$ volume percent ratio in the acid gas feed stream averaged 68.9/31.1 and Claus plant sulfur recovery efficiency ranged from 95.20 percent to 97.81 percent, whereas liquid sulfur production ranged from 230.3 Mg/d (226.7 LT/D) to 164.6 Mg/d (162 LT/D). Stack gas temperature was an average 859 K (1086°F). Sulfur recovery efficiency is equal to sulfur production divided by the sum of sulfur production and sulfur in the Claus tail gas.

### C.1.6 Shell Oil's Person Plant Facility[6]

The Person Plant facility is a Claus sulfur recovery facility that utilizes two parallel two-stage Claus trains with a common third-stage reactor bed. The plant began operations in 1962 as two-stage unit. A

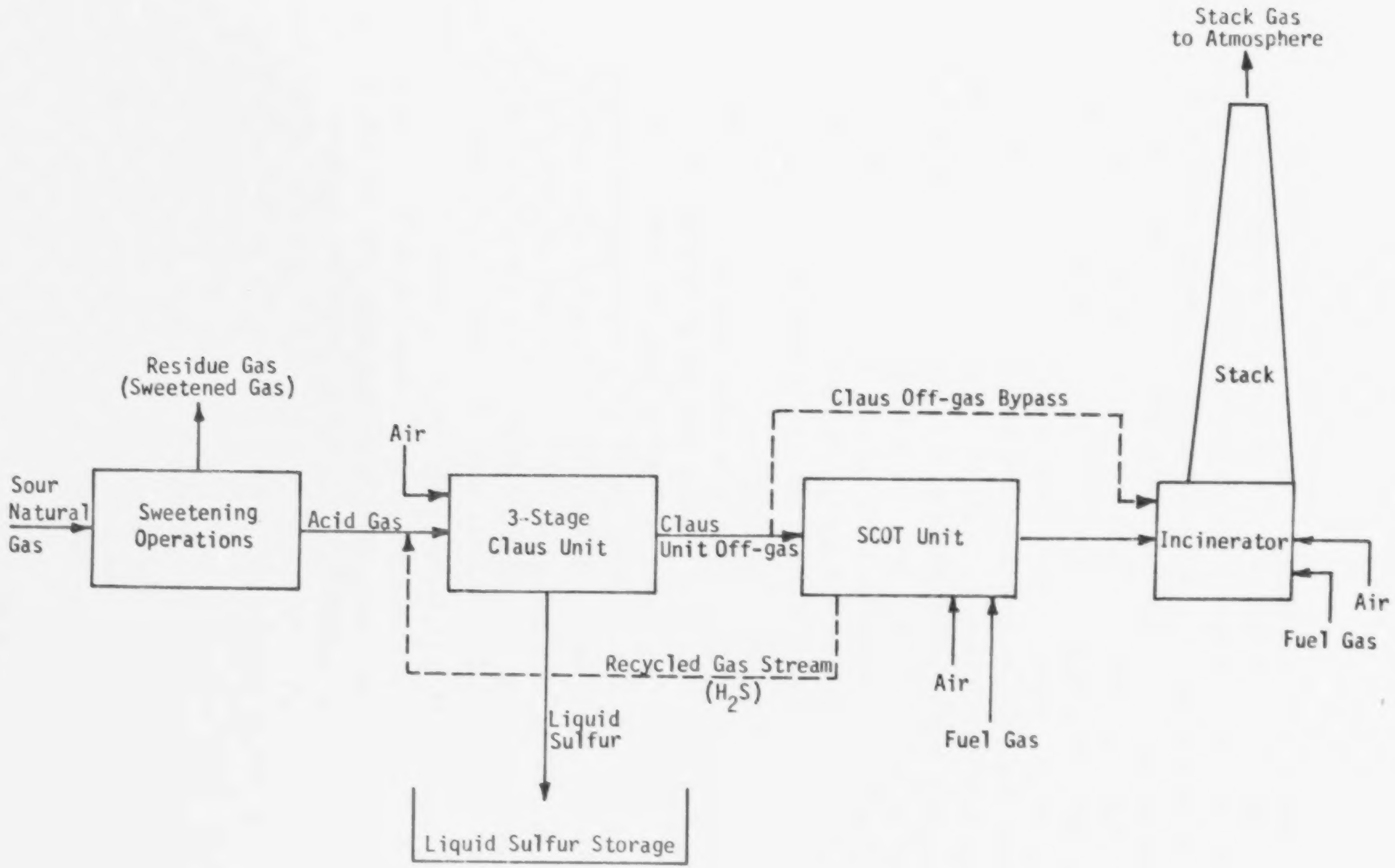
Stack Gas to Atmosphere
Stack
Residue Gas (Sweetened Gas)
Air
Claus Off-gas Bypass
Sour Natural Gas
Sweetening Operations
Acid Gas
3-Stage Claus Unit
Claus Unit Off-gas
SCOT Unit
Incinerator
Air
Fuel Gas
Recycled Gas Stream ($H_2S$)
Air
Liquid Sulfur
Fuel Gas
Liquid Sulfur Storage

Figure C-7. Simplified flow diagram for Exxon's Blackjack Creek Facility.

Table C-20. EXXON'S BLACKJACK CREEK FACILITY ANNUAL STACK EMISSION TESTS DATA[a]

| Test date | Acid gas volumetric flow rate entering the Claus unit (includes recycle flow) |  | Total sulfur entering the Claus unit (includes recycle sulfur) |  | Net inlet sulfur to the Claus unit |  | Stack $SO_2$ emissions |  | $H_2S/CO_2$ ratio in the acid gas entering the Claus unit | Sulfur recovery efficiency, % |  |
|---|---|---|---|---|---|---|---|---|---|---|---|
|  | NM³/s | MMscf/day | Mg/d | lb/day | Mg/d | lb/day | kg/d | lb/day |  | With SCOT unit | Without SCOT unit |
| 3/07/80 | 1.134 | 3.459 | 108.2 | 238,540 | 105.4 | 232,458 | 412 | 908 | 87.5/12.5 | 99.81 | 97.26 |
| 4/19/79 | 1.194 | 3.644 | 108.6 | 239,472 | 105.2 | 231,900 | 422 | 931 | 85.8/14.2 | 99.80 | 96.64 |
| 4/14/78 | 1.110 | 3.386 | 101.4 | 223,595 | 97.9 | 215,886 | 161 | 354 | 86.7/13.3 | 99.92 | 96.47 |
| 7/20/77 | 1.090 | 3.326 | 101.4 | 223,568 | 97.6 | 215,217 | 216 | 476 | 85.6/14.4 | 99.89 | 96.16 |

[a] Supplied by the facility.

Table C-21. SHELL OIL'S BRYANS MILL FACILITY SOURCE EMISSION TESTS DATA[a]

| Test date | Liquid sulfur production | | [b]$SO_2$ emissions | | Sulfur recovery efficiency, % | $H_2S/CO_2$ volume percent ratio | Incinerator temperature | |
|---|---|---|---|---|---|---|---|---|
| | Mg/d | LT/D | Mg/d | LT/D | | | K | °F |
| 6/11/75 | 212.6 | 209.28 | 19.5 | 19.18 | 95.60 | 69.3/28.6 | 887 | 1,137 |
| 12/10/74 | 207.6 | 204.34 | 14.2 | 14.00 | 96.69 | 70.2/28.3 | 872 | 1,110 |
| 10/10/74 | 207.5 | 204.20 | 13.0 | 12.76 | 96.97 | | 862 | 1,092 |
| 6/13/74 | 196.8 | 193.70 | 18.2 | 17.92 | 95.60 | | | |
| 6/05/74 | 211.8 | 208.42 | 17.4 | 17.13 | 96.05 | | | |
| 4/23/74 | 223.6 | 220.03 | 12.1 | 11.92 | 97.40 | | | |
| 4/17/74 | 205.4 | 202.20 | 17.6 | 17.32 | 95.90 | | | |
| 12/18/73 | 181.2 | 178.30 | 8.1 | 7.98 | 97.81 | | | |
| 10/09/73 | 178.9 | 176.04 | 16.4 | 16.18 | 95.60 | | | |
| 10/08/73 | 199.3 | 196.20 | 15.2 | 14.96 | 96.30 | | | |
| 7/11/73 | 198.6 | 195.50 | 20.0 | 19.71 | 95.20 | 67.7/29.8 | 826 | 1,027 |
| 3/29/73 | 164.6 | 162.00 | 12.6 | 12.38 | 96.32 | 68.4/30.4 | | |
| 3/28/73 | 182.4 | 179.53 | 14.4 | 14.18 | 96.20 | | | |
| 3/27/73 | 191.4 | 188.37 | 13.9 | 13.66 | 96.50 | | | |
| 3/26/73 | 208.2 | 204.93 | 18.1 | 17.79 | 95.84 | | | |
| 12/20/72 | 177.7 | 174.88 | 10.9 | 10.76 | 97.02 | | 853 | 1,075 |
| 12/19/72 | 194.8 | 191.73 | 11.3 | 11.16 | 97.17 | | 853 | 1,075 |
| 9/22/72 | 230.3 | 226.68 | 15.0 | 14.74 | 96.85 | | | |
| 8/17/72 | 184.4 | 181.44 | 14.7 | 14.50 | 96.16 | | | |
| 3/22/72 | 222.6 | 219.10 | 12.1 | 11.86 | 97.36 | | | |

[a] Supplied by the facility.

[b] Includes TRS.

second two-stage train was placed in operation in 1965. Third-stage converter common to both the trains along with Claus plant tail gas incinerator were added in 1970-71 revisions. During 1975, the average sour natural gas flow rate was 11.2 $Nm^3/s$ (34.1 MMscf/day), and the average daily sulfur production was 13.9 Mg/d (13.7 LT/D). The $H_2S$ concentration in the sour natural gas averaged 1.12 percent by volume, with an average acid gas feed stream $H_2S/CO_2$ volume percent ratio of 20.6/79.4. Stack gas temperature averaged 822 K (1020°F). A summary of the source emission tests conducted during 1973 through 1975 is presented in Table C-22. Sulfur production ranged from 14.4 Mg/d (14.2 LT/D) to 23.2 Mg/d (22.8 LT/D) and sulfur recovery efficiency ranged from 92.95 percent to 97.7 percent. Sulfur recovery efficiency is equal to sulfur production divided by the total of sulfur production and sulfur in the Claus tail gas.

### C.1.7 Exxon's Santa Rosa Facility[7]

The Santa Rosa facility is a three-stage Claus sulfur recovery facility processing 8.4 $Nm^3/s$ (25.7 MMscf/day) sour natural gas. The facility produced 104.7 Mg/d liquid sulfur (103 LT/D). Acid gas flow rate was 1.2 $Nm^3/s$ (3.65 MMscf/day) with 77.3 percent by volume $H_2S$ in it. The source emission tests conducted by EPA in June 1975 indicated sulfur recovery efficiency of 97.7 percent achieved with 80.1 percent by volume $H_2S$ in the acid gas feed stream (typical sulfur recovery efficiency range 96 percent to 97 percent). The incinerator was operated at temperatures of 783 K (950°F), 839 K (1050°F) and 950 K (1250°F). Total reduced sulfur (COS, $CS_2$ and $H_2S$) was reduced from 219 ppmv to 18 ppmv as the incinerator temperature was increased from 783 K (950°F) to 950 K (1250°F).

### C.1.8 Exxon's Flomaton Facility[8]

The Flomaton facility is a three-stage Claus sulfur recovery facility designed to process 11.5 $Nm^3/s$ (35 MMscf/day) sour natural gas with sulfur recovery of 135.8 Mg/d (133.7 LT/D). The acid gas flow rate is 6.1 $Nm^3/s$ (18.5 MMscf/day) acid gas and contains 20.6 percent by volume $H_2S$ in it. During the source emission test program in November 1973, the facility recovered 130.1 Mg/d (128 LT/D) liquid sulfur with 96.7 percent sulfur recovery efficiency.

## Table C-22. SHELL OIL'S PERSON PLANT FACILITY SOURCE EMISSION TESTS DATA[a]

| Test date | Liquid sulfur production Mg/d | LT/D | $SO_2$ emissions[b] Mg/d | LT/D | Sulfur recovery efficiency, % |
|---|---|---|---|---|---|
| 11/06/75 | 21.4 | 21.08 | 3.3 | 3.196 | 92.95 |
| 3/10/75 | 14.4 | 14.16 | 2.0 | 1.926 | 93.63 |
| 4/10/74 | 18.5 | 18.24 | 1.3 | 1.252 | 96.68 |
| 4/18/73 | 23.2 | 22.82 | 1.1 | 1.066 | 97.70 |

[a] Supplied by the facility.

[b] Includes TRS.

### C.1.9 Aquitaine of Canada, Ltd.'s Ram River Facility[9]

At the Ram River facility, acid gas containing 84 percent $H_2S$ by volume is processed by four two-stage Claus sulfur recovery units operating in parallel. Each of the four Claus trains is designed to recover 1,016 Mg/d liquid sulfur (1,000 LT/D). The tail gases from these four Claus plants are routed to two Sulfreen tail gas cleanup units. Each Sulfreen unit consists of three reactors, with two in adsorption cycle and one in regeneration cycle at any given time. This is the largest sulfur recovery facility of its kind in Canada. During the source emission tests conducted by U.S. EPA in September 1974, sulfur recovery efficiency ranged from 97.4 percent to 98.3 percent (average of 98 percent) while sulfur intake ranging from 3568 Mg/d (3512 LT/D) to 4100 Mg/d (4036 LT/D).

### C.1.10 Chevron Standard, Ltd.'s Fox Creek Facility[10]

The fox Creek facility consists of two four-stage Claus sulfur recovery units, each with a capacity of recovering 1727 Mg/d (1700 LT/D) liquid sulfur. This is the largest facility of its kind in Canada. The acid gas feed stream containing 77 percent $H_2S$ by volume flows into each Claus unit at an average rate of 19.7 $Nm^3/s$ (60 MMscf/day). During the source emission tests conducted by U.S. EPA in November 1974, sulfur production from each Claus unit ranged from 1784 Mg/d (1756 LT/D) to 1814 Mg/d (1785 LT/D), with sulfur recovery efficiency ranging from 98.6 percent to 98.7 percent.

## C.2 REFERENCES

1. EMB Report No. 80-OSP-4. Onshore Production of Crude Oil and Natural Gas - Sulfur Plants, Emission Test Report, $SO_2$ testing at the Warren Petroleum Monument Plant, Monument, New Mexico. June 1981. Docket No. A-80-20-A, Entry II-A-15.

2. EMB Report No. 80-OSP-9. Onshore Production of Crude Oil and Natural Gas - Sulfur Plants, Emission Test Report, $SO_2$ testing at the Getty Oil New Hope Plant, New Hope, Texas. July 1981. Docket No. A-80-20-A, Entry II-A-14.

3. EMB Report No. 80-OSP-6. Onshore Production of Crude Oil and Natural Gas - Sulfur Plants, Emission Test Report, $SO_2$ testing at the Shell Oil Thomasville Plant, Thomasville, Mississippi. July 1981. Docket No. A-80-20-A, Entry II-A-17.

4. Annual stack test reports (1980, 1979, 1978, and 1977) on the Blackjack Creek facility located in the Jay Field, Florida, submitted to Florida Department of Environmental Regulation by Exxon Company, U.S.A. New Orleans, Louisiana. Docket No. A-80-20-A, Entry II-D-17.

5. Bryans Mill plant (Cass County, Texas). Process data and general information (operation and test data for 1975, 1974, 1973, and 1972) presented to EPA by Shell Oil Company in January 1976. Docket No. A-80-20-A, Entry II-D-11.

6. Person plant (Karnes County, Texas) Gas treating and sulfur plant data (operation and test data for 1975, 1974, and 1973) for EPA by Shell Oil Company in January 1976. Docket No. A-80-20-A, Entry II-D-11.

7. EMB Report No. 75-SRY-9. Air Pollution Emission Test, Emissions from an oil and natural gas field sulfur recovery plant at Exxon Company, U.S.A., Santa Rosa plant, Jay, Florida, February 1976. U.S. EPA, Office of Air Quality Planning and Standards, Emission Measurement Branch, Research Triangle Park, N.C. and additional information on the Santa Rosa facility received from EPA. Docket No. A-80-20-A, Entry II-A-7.

8. Sulfur Recovery Unit Performance Test, November 12, 1973, on Flomaton production facility, submitted to State of Alabama, Air Pollution Control Commission by Exxon Company, U.S.A., New Orleans, Louisiana, and composition, flow data on the Flomaton facility received from EPA. Docket No. A-80-20-A, Entry II-A-21.

9. Air Pollution Emission Test, Report No. 75-SRY-6. Source testing of Ram River plant sulfur recovery facility. Aquitaine of Canada, Ltd., Ram River, Alberta. November 1975. U.S. EPA, Office of Air Quality Planning and Standards, Emission Measurement Branch, Research Triangle Park, N.C. Docket No. A-80-20-A, Entry II-A-6.

10. Air Pollution Emission Test, Report No. 75-SRY-5. Source testing of Fox Creek plant sulfur recovery facility. Chevron Standard, Ltd., Fox Creek, Alberta. November 6-8, 1974. U.S. EPA, Office of Air Quality Planning and Standards, Emission Measurement Branch, Research Triangle Park, N.C. Docket No. A-80-20-A, Entry II-A-5.

# APPENDIX D - EMISSION MEASUREMENT AND CONTINUOUS MONITORING

## D.1 EMISSION MEASUREMENT METHODS

During the standard support study for sulfur recovery plants at crude oil and natural gas processing facilities, three facilities were tested. These facilities were equipped with Claus sulfur recovery plants followed by tail-gas incinerators fired with sweet natural gas.

The test program at these facilities consisted of measurement of the sulfur recovery efficiency of the sulfur recovery plant and the nitrogen oxides emission rate. The sulfur omission rate was determined by sampling the incinerator exhaust stack to measure sulfur dioxide, hydrogen sulfide, and total reduced sulfur. EPA Reference Methods 6 and 11 (40 FR Part 60, Appendix A) and proposed Method 16A (Federal Register, Volume 46, No. 117, June 18, 1981), respectively, were used successfully without modification. The exhaust gas flow rate was determined by Methods 1, 2, 3, and 4. For Method 3, the gas composition was analyzed using thermal conductivity gas chromatography instead of Orsat apparatus.

The incinerator exhaust sulfur rate was combined with the sulfur recovery rates determined from plant instrumentation to calculate sulfur recovery efficiency.

Measurements of nitrogen oxides were performed using EPA Method 7.

All gaseous measurements were conducted at a single point in the exhaust stacks because the sampling locations at these facilities were approximately 8-10 equivalent stack diameters downstream of disturbances, and the velocity profiles were relatively uniform. In addition, no glass wool plugs were used for filtration in the gaseous sampling trains because of the low particulate concentrations in the exhaust gases.

In previous testing programs at similar facilities in petroleum refineries, the sulfur species in the emissions were measured by a different procedure. These facilities were equipped with Claus sulfur recovery units and additionally with tail gas treatment units. The full test programs are described in Appendix C of the "Standard Support and Environmental Impact Statement, Volume I: Proposed Standards of Performance for Petroleum Refinery Sulfur Recovery Plants," EPA Publication No. EPA-450/2-76-016a, September 1976. EPA Method 15 (40 CFR Part 60, Appendix A), identified as Method 18 in the above document, was used to determine individual reduced sulfur compounds. This technique employs a field gas chromatograph equipped with a flame photometric detector. The use of a chromatographic technique instead of a wet chemical approach, as is in proposed Method 16A, was advantageous because the total sulfur concentrations were lower and one of the purposes of testing was to identify the species of sulfur.

A potential limitation on the use of Method 16A at plants equipped with a reduction-type tail gas treatment device would be that insufficient oxygen is available in the exhaust gases to support complete oxidation of reduced sulfur to sulfur dioxide prior to the sample collection in the impinger train. At those locations, it will be necessary to use Method 15.

## D.2 PERFORMANCE TEST METHODS.

The EPA reference test methods are available for all measurements necessary to determine the total sulfur emissions from sulfur recovery plants. Process instrumentation is available to measure the additional parameters necessary to compute the sulfur recovery efficiency achieved by a system, and to compute the minimum required emission reduction under the regulation.

The format of the draft sulfur regulation is in terms of a minimum $SO_2$ emission reduction efficiency. In addition, this minimum reduction is a function of the total sulfur feed rate to the sulfur recovery process and the concentration of hydrogen sulfide in the feed. In order to determine the emission reduction limitation required, these parameters must be measured during a performance test.

The procedures specified for the measurement of $H_2S$ concentration in the acid gas feed are the Tutwiler(1) procedure and process chromatography. It is expected that all new facilities will use one of these techniques for routine process monitoring. The Tutwiler procedure is a direct iodine titration of an $H_2S$ sample in a specially designed burette. If chromatographic techniques are used, then the ASTM Recommended Practices for General Gas Chromatography Procedures(2) should be followed. It is required that at least one sample be collected and analyzed each hour during the performance test period, which is a minimum of 12 hours. The concentration is in terms of mole percent.

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(1) "Gas Engineer's Handbook," Fuel Gas Engineering Practices, First Edition, The Industrial Press, New York 1966, p. 6-25. Docket No. A-80-20-A, Entry II-I-67.

(2) ASTM Method E-260, 1973.

All the results are averaged to determine a mean $H_2S$ concentration for the performance test for use in the regulation equation.

The sulfur feed rate to the recovery unit is determined by multiplying the average $H_2S$ content by the volumetric flow rate of the acid gas stream. Process flow meters are available to measure this flow rate and are expected to be available at new facilities. The meter type is not specified by the regulation, but it should be installed and operated according to the manufacturer's recommendations and accepted standard practices. The volumetric flow is expressed in terms of standard cubic feet per day. Standard conditions are 20°C and 760 mm Hg. The average flow rate for the performance test period is used with the average $H_2S$ content to calculate the average sulfur feed rate in terms of long tons per day.

The average sulfur feed rate and the average $H_2S$ content of the acid gas are used in the regulation equation to determine the minimum required $SO_2$ reduction efficiency.

The actual $SO_2$ reduction efficiency achieved during the performance test is determined by measurements of the stack emissions and the sulfur production rate. The ratio of the sulfur accumulated in the product storage tank to the total sulfur out of the process is used as the measure of actual recovery efficiency achieved.

The sulfur production rate is determined by measuring the accumulation of liquid sulfur in the product tanks.

Process level indicators or manual soundings are used to determine the level of liquid sulfur in the storage tank at the beginning and end of each test run. Each test run is a minimum of 4 hours, and triplicate runs are required. The level readings are converted to mass of sulfur using the tank geometry calibrations and the density of the liquid product. The change in mass is divided by the test run duration to obtain a sulfur production rate. The individual results from the three runs are averaged to obtain a mean value for use in the actual recovery efficiency equation.

The determination of the sulfur emission rate requires the measurement of sulfur dioxide, total reduced sulfur compounds, and exhaust gas flow rate. The exhaust gas flow rate is determined using EPA Methods 1, 2, 3, and 4, for traverse point selection, velocity measurement, gas molecular weight, and moisture content, respectively. These procedures are applicable to all sources.

The procedures used to determine the total sulfur compound concentration will vary depending on the type control device. The EPA Method 6 is used to determine $SO_2$ concentration at all facilities. For those facilities where final sulfur removal is by an oxidation process, or where there is more than 1.0 percent by volume of oxygen in the exhaust gas, EPA Method 16A may be used to measure the total reduced sulfur compound concentration. The use of this method is limited to those sources where enough oxygen is present in the sample gas to allow oxidation of all reduced compounds to $SO_2$ in a furnace used in the procedure. For those control devices that incorporate a reduction atmosphere where little excess oxygen is present in the exhaust gas, EPA Method 15 is used to measure the reduced sulfur compound concentration.

The testing period for each run is a minimum of 4 hours. For Methods 6 and 16A, the run may be one 4-hour sample, or a series of shorter samples whose total duration is 4 hours. If multiple samples are used, then a time-weighted mean is computed for the run result. For Method 15, 16 samples spaced equally over at least a 4-hour period are required for each run. The average for the 16 samples is computed as a run result.

Since the stack velocity, gas molecular weight, and moisture content are not expected to vary significantly during the run, the minimum sampling times per run are less. A velocity traverse is required at the beginning and end of each run. A moisture sample of at least 10 minutes each is required at the beginning and end of each run. The gas sample is either integrated continuously over the sample run, or grab samples may be collected and analyzed at 1-hour intervals. The mean results of these measurements are used to calculate a stack volumetric flow rate for each run.

Finally, for each run the total sulfur emission rate is calculated by adding the $SO_2$ and TRS concentrations (converted to $SO_2$ equivalents), multiplying by the stack gas flow rate, and converting the mass result to an elemental sulfur basis.

The sulfur emission rate for each run is averaged for use in the actual sulfur recovery rate equation.

The measured sulfur emission rate and sulfur production rate are combined in the equation:

$$R = \frac{S}{S + E} (100)$$

where: R = the actual sulfur recovery rate achieved during the performance test (%)

S = the average sulfur production rate during the performance test, Kg/hr as sulfur

E = the average total sulfur emission rate during the performance test, Kg/hr as sulfur

The minimum required $SO_2$ emission reduction efficiency is calculated by:

$$Z = 88.51\ X^{0.0101}\ Y^{0.0125}$$

where: Z = the minimum required $SO_2$ emission reduction efficiency

X = the average sulfur feed rate to the sulfur recovery unit during the performance test (LT/D)

Y = the average $H_2S$ concentration in the acid gas feed during the performance test (mole percent)

To determine compliance, R is compared to Z. If R is greater than or equal to Z, then the facility is in compliance.

The cost of a performance test using these procedures is estimated to be from $10,000 to $15,000.

## D.3 CONTINUOUS MONITORING

The draft standard requires continuous emission monitoring of the sulfur mass rate and monitoring devices to measure and record the rate of liquid sulfur production and the temperature of the gas leaving the combustion zone of a combustion device that is a continually operated part of the sulfur recovery system.

The monitoring device for temperature is expected to be a standard thermocouple or similar device. Such devices when selected and installed according to standard accepted practices are capable of measuring within ±1 percent, or within ±10°F at 1000°F.

The monitoring device for the rate of liquid sulfur accumulation is expected to be a level indicator in the product storage tank. The resolution necessary to obtain a ±2 percent accuracy for this level indicator will depend on the geometry of the storage tank and the relative accumulative rate at each facility. The alternative of measuring the volumetric flow rate of liquid sulfur would also be acceptable.

The continuous monitoring system necessary to measure the sulfur mass emission rate would consist of an instrument to measure the sulfur species concentration, an instrument to measure the stack gas velocity or volumetric flow rate, and a data processing device to combine the measurements into a mass rate output.

For those sources that incorporate a final flue gas oxidation in the process unit, an $SO_2$ analyzer would be used to measure sulfur concentration. Performance Specification 2 would be used to evaluate the instrument performance. Method 6 would be used to evaluate the relative accuracy of the analyzer.

If the sulfur recovery system does not incorporate a continually operated oxidation device, and the sulfur species emitted are reduced compounds, then a total reduced sulfur analyzer would be used to measure total sulfur emissions. Performance Specification 5 would be used to evaluate the instrument system, with EPA Method 15 used to determine relative accuracy.

The EPA has not proposed a performance specification for velocity or volumetric flow measurement devices. If the volumetric flow device is considered to be an emission monitoring system, then these would be necessary before affected facilities would be required to install and operate these systems for continuous monitoring purposes. However, if the exhaust gas flow measurement is considered to be a process parameter monitoring device, then the precedent has been that manufacturers' specifications and nationally accepted recommended standard practices may be cited as selection, installation, and operation specifications.

Instrument systems are commercially available that are designed to measure $SO_2$ or TRS based total sulfur mass emission rates. At this time, EPA has not conducted field evaluations of these systems, and the only information available as to the accuracy, operability, and reliability of these systems are vendor information and literature sources.

The estimated cost of the monitoring devices for temperature and level indicator or product flow rate are about $3,000 each installed. However, these devices are usually installed as standard equipment and any additional cost attributable to the regulation would be a different data recording system.

The estimated costs of a sulfur mass monitoring system are an installed capital cost of $30,000 to $40,000 and an annual operating cost of $10,000 to $15,000/year.

# SULFUR RECOVERY STUDY

## ONSHORE SOUR GAS PRODUCTION FACILITIES

APPENDIX E

**July 1981**

# SECTION 1 (SECTION E-1)

# DESIGN BASIS

## INTRODUCTION

This study was prepared for the development of New Source Performance Standards for $SO_2$ emissions from Onshore Natural Gas Production Facilities. This report provides investment costs, direct operating cost data, process descriptions, process flow diagrams, and atmospheric emissions for 39 cases with different sizes and combinations of sulfur recovery and tail gas processes. For explanations of process names and abbreviations used, refer to Definitions in this section.

## DESIGN CRITERIA AND ASSUMPTIONS

Listed below are assumptions used for design criteria and cost estimating for this study:

(1) Barometric pressure is 14.7 psia.

(2) Acid gases are available at 100°F and 24.7 psia, saturated with water, and contain 0.5 vol % methane (wet basis).

(3) Steam is produced at 250, 50, and 15 psig from the Claus units.

(4) Thermal oxidizers are operated at 1,200°F with 25% excess air (to reduce $H_2S$ content below 10 ppmv).

(5) For cases with waste heat boilers associated with thermal oxidizers, steam is produced at 250 psig for the 100-LT/D cases and at 600 and 250 psig for the 555- or 1,000-LT/D cases.

(6) Treated, deaerated boiler feedwater is available at 320 psig and 230°F.

(7) Cooling water is available at 85°F and is returned at 110°F.

(8) Stack heights vary from 100 to 600 ft depending on the quantity of sulfur dioxide emissions (see Tables 1-1, 1-2, and 1-3); stack height was set to achieve approximately uniform levels of ground-level $SO_2$ concentration.

(9) Investment costs are based on January 1981 Gulf Coast prices.

(10) In the BSR (hydrogenation) sections, steam is produced in the 100-LT/D units at 50 psig and in the 555- or 1,000-LT/D units at 450 and 50 psig.

(11) For the cases with 1,000-LT/D sulfur input feeding acid gas with a 50/50 $H_2S/CO_2$ ratio, the plants are approximately the maximum economical size and weaker gases would require building two trains. In order to keep all cases in only one train, the maximum sulfur input with 20/80 $H_2S/CO_2$ acid gas feed is 555 LT/D. These plants are about the same physical size as the 1,000-LT/D plants with 50/50 $H_2S/CO_2$ acid gas feed.

## DEFINITIONS

The following list provides definitions, terminology, and abbreviations used in this report.

Acid Gas: The gas containing $H_2S$ (and $CO_2$) resulting from treating of sour natural gas. Also contains small amounts of hydrocarbons and water.

BSR: The hydrogenation/hydrolysis and cooling section for converting all sulfur compounds to $H_2S$ for further treatment as described in the process description sections for the BSR/MDEA process, Beavon Sulfur Removal Process (BSRP), and the BSR/Selectox process. Hydrogen required for the hydrogenation reaction is generated within the process.

Claus: The name of the gas phase reaction where 2 mols of $H_2S$ combine with 1 mol of $SO_2$ to form 3 mols of sulfur. In what has come to be called the Claus sulfur recovery process, a reaction furnace is included for combustion to provide the necessary one-third of $SO_2$ for the reaction.

End-of-Run: After a certain period of operation, the activity of such catalysts as alumina used in the Claus process declines to a point where it is economical to shut down and replace it with fresh catalyst, say after 2 years. The $SO_2$ emissions in the tables in this report are based on end-of-run recoveries. The LT/D of sulfur recovered in Tables 6-9, 6-10, and 6-11 are average during run, calculated by averaging the start-of-run and end-of-run recoveries. These average sulfur recoveries can then be used to calculate credits for sulfur sales against operating costs.

Gas Components:

| | |
|---|---|
| $H_2S$ | Hydrogen Sulfide |
| $SO_2$ | Sulfur Dioxide |
| $O_2$ | Oxygen |
| $N_2$ | Nitrogen |
| CO | Carbon Monoxide |
| $CO_2$ | Carbon Dioxide |
| COS | Carbonyl Sulfide |
| $CS_2$ | Carbon Disulfide |
| $H_2O$ | Water |

Lb Mols/Hr: The pounds per hour divided by the molecular weight of the component.

LT/D: Long tons (2,240 pounds) per day.

MDEA: Methyl diethanolamine, used in aqueous solution to absorb $H_2S$ and part of the $CO_2$ from gas.

MM: Millions.

SCOT: A process proprietary to Shell Oil Company. It is believed similar to BSR/MDEA, and published literature indicates use of extraneous hydrogen for the hydrogenation reaction.

Selectox: A catalyst developed by Union Oil Company of California which oxidizes $H_2S$ directly to elemental sulfur and in a temperature range substantially lower than required for uncatalyzed oxidation.

Sour Gas: Natural gas containing $H_2S$. When the $H_2S$ and usually the $CO_2$ are absorbed and stripped from the hydrocarbons in any of various treating units, the stripped gas containing the $H_2S$ and $CO_2$ is called acid gas.

Thermal Oxidizer: A combustion chamber for converting all sulfur compounds in the tail gas to $SO_2$ by burning the required amount of fuel. These are sometimes called incinerators.

Vol. ppm or ppmv: The parts per million of a component on a volume basis.

## FEED GAS COMPOSITIONS

The feed gas (acid gas) compositions in pound mols per hour for all the cases are provided in Table 1-1. In all of the cases the acid gas is saturated with water at 100°F and 24.7 psia and contains 0.5 vol % of methane. These conditions were assumed for this study for the sake of consistency. Actual conditions will, of course, vary somewhat but will have a negligible effect on the results and conclusions. Most of the solutions used to extract acid gas from natural gas contain an amine in water, resulting in water-saturated acid gas; some physical solvents (e.g., Selexol) produce dry acid gas.

## STACK HEIGHTS

The sulfur emissions as $SO_2$ in vol % and in lb/hr are provided for each of the cases, along with the assumed stack heights, in Tables 1-2, 1-3, and 1-4.

These stack heights were determined from experience to roughly approximate the same ground-level concentration of $SO_2$ for all cases.

In the various tables under "Thermal Oxidizers", the BSRP cases state No* with a footnote: *Combustor only. At the top of the Stretford section absorber, there is a combustor for emergency use only with a stack above the combustor. The elevation at the top of this stack varies from about 100 to 150 ft above grade depending on the size and location of the unit.

The sulfur emissions are end of run. The vol % concentrations can be converted to parts per million (ppmv) by multiplying by 10,000. The lb/hr of $SO_2$ can be converted to LT/D of $SO_2$ by multiplying by 0.0107.

Table 1-1 (Table E-1)

## Onshore Sour Natural Gas Sulfur Recovery Study for TRW

### Feed Gas Compositions

| $H_2S/CO_2$ Ratio | 80/20 | 50/50 | | | |
|---|---|---|---|---|---|
| Sulfur Input, LT/D | 1,000 | 5 | 10 | 100 | 1,000 |
| Case No. | 19 | 1,3 | 5,7 | 9,11,13<br>15 & 17 | 20,23,26<br>29 & 32 |
| Mols/hr | | | | | |
| $H_2S$ | 2,911.2 | 14.56 | 29.11 | 291.1 | 2,911.2 |
| $CO_2$ | 727.8 | 14.56 | 29.11 | 291.1 | 2,911.2 |
| $CH_4$ | 19.0 | .15 | .30 | 3.0 | 30.4 |
| $H_2O$ | 146.3 | 1.17 | 2.34 | 23.4 | 234.1 |
| Total | 3,804.3 | 30.44 | 60.86 | 608.6 | 6,086.9 |

| $H_2S/CO_2$ Ratio | 20/80 | | | | 12.5/87.5 | | |
|---|---|---|---|---|---|---|---|
| Sulfur Input, LT/D | 5 | 10 | 100 | 555 | 5 | 10 | 100 |
| Case No. | 2,4 | 6,6B,8 & 8B | 10, 10B, 12,14, 16, 18 | 21,24 27, 30,33 | 4A | 6A,6C, 8A & 8C | 10A & 10C |
| Mols/hr | | | | | | | |
| $H_2S$ | 14.56 | 29.11 | 291.1 | 1,615.7 | 14.56 | 29.11 | 291.1 |
| $CO_2$ | 58.24 | 116.44 | 1,164.4 | 6,642.8 | 101.92 | 203.77 | 2,037.7 |
| $CH_4$ | .38 | .76 | 7.6 | 42.2 | .61 | 1.22 | 12.2 |
| $H_2O$ | 2.93 | 5.85 | 58.5 | 324.8 | 4.68 | 9.36 | 93.6 |
| Total | 76.11 | 152.16 | 1,521.6 | 8,445.5 | 121.77 | 243.46 | 2,350.4 |

Table 1-2 (Table E-2)

Onshore Sour Natural Gas Sulfur Recovery Study

End-of-Run Sulfur Emission as $SO_2$ and Stack Height

Acid Gas $H_2S/CO_2$ Ratios of 80/20 and 50/50

| CASE NO. | SULFUR INPUT (LT/D) | SULFUR RECOVERY UNIT |  | TAIL GAS UNIT | THERMAL OXIDIZER |  | SULFUR AS $SO_2$ |  | STACK |  |
|---|---|---|---|---|---|---|---|---|---|---|
|  |  | NO. STAGES | TYPE |  | INCLUDED | WASTE HEAT RECOVERY | VOL % | LB/HR | HEIGHT (FT) | TEMP (°F) |
| **Acid Gas $H_2S/CO_2$ Ratio - 80/20** |  |  |  |  |  |  |  |  |  |  |
| 19 | 1,000 | 3 |  | No | Yes | Yes | 1.270 | 7514 | 550 | 450 |
| **Acid Gas $H_2S/CO_2$ Ratio - 50/50** |  |  |  |  |  |  |  |  |  |  |
| 1 | 5 | - | No | No | Yes | No | 4.964 | 933 | 250 | 1500 |
| 3 | 5 | 2 | Recycle Selectox | No | Yes | No | 1.293 | 70 | 100 | 1200 |
| 5 | 10 | 2 | Claus | No | Yes | No | 1.268 | 138 | 100 | 1200 |
| 7 | 10 | 3 | Claus | No | Yes | No | 0.817 | 88 | 100 | 1200 |
| 9 | 100 | 3 | Claus | No | Yes | Yes | 0.817 | 884 | 225 | 450 |
| 11 | 100 | 3 | Claus | BSR/MDEA | Yes | Yes | 0.011 | 12 | 100 | 450 |
| 13 | 100 | 3 | Claus | BSRP | No* | No | 0.004 | 3 | - | 100 |
| 15 | 100 | 3 | Claus | Sulfreen | Yes | Yes | 0.329 | 354 | 150 | 450 |
| 17 | 100 | 2 | Claus | BSR/Selectox | Yes | Yes | 0.350 | 339 | 150 | 450 |
| 20 | 1,000 | 3 | Claus | No | Yes | Yes | 0.817 | 8847 | 600 | 450 |
| 23 | 1,000 | 3 | Claus | BSR/MDEA | Yes | Yes | 0.011 | 115 | 100 | 450 |
| 26 | 1,000 | 3 | Claus | BSRP | No* | No | 0.004 | 23 | - | 100 |
| 29 | 1,000 | 3 | Claus | Sulfreen | Yes | Yes | 0.330 | 3539 | 400 | 450 |
| 32 | 1,000 | 2 | Claus | BSR/Selectox | Yes | Yes | 0.350 | 3389 | 400 | 450 |

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*Combustor only

Table 1-3 (Table E-3)

**Onshore Sour Natural Gas Sulfur Recovery Study**

**End-of-Run Sulfur Emission as $SO_2$ and Stack Height**

**Acid Gas $H_2S/CO_2$ Ratio of 20/80**

| CASE NO. | SULFUR INPUT (LT/D) | SULFUR RECOVERY UNIT |  | TAIL GAS UNIT | THERMAL OXIDIZER |  | SULFUR AS $SO_2$ |  | STACK |  |
|---|---|---|---|---|---|---|---|---|---|---|
|  |  | NO. STAGES | TYPE |  | INCLUDED | WASTE HEAT RECOVERY | VOL % | LB/HR | HEIGHT (FT) | TEMP (°F) |
| 2 | 5 | - | No | No | Yes | No | 5.34 | 933 | 250 | 1500 |
| 4 | 5 | 2 | Recycle Selectox | No | Yes | No | 0.851 | 90 | 100 | 1200 |
| 6 | 10 | 2 | Claus | No | Yes | No | 0.882 | 184 | 125 | 1200 |
| 6B | 10 | 2 | Recycle Selectox | No | Yes | No | 0.851 | 179 | 125 | 1200 |
| 8 | 10 | 3 | Claus | No | Yes | No | 0.591 | 123 | 100 | 1200 |
| 8B | 10 | 3 | Recycle Selectox | No | Yes | No | 0.491 | 103 | 100 | 1200 |
| 10 | 100 | 3 | Claus | No | Yes | Yes | 0.591 | 1230 | 275 | 450 |
| 10B | 100 | 3 | Recycle Selectox | No | Yes | Yes | 0.491 | 1034 | 250 | 450 |
| 12 | 100 | 3 | Claus | BSR/MDEA | Yes | Yes | 0.009 | 22 | 100 | 450 |
| 14 | 100 | 3 | Claus | BSRP | No* | No | 0.007 | 9 | - | 100 |
| 16 | 100 | 3 | Claus | Sulfreen | Yes | Yes | 0.238 | 493 | 175 | 450 |
| 18 | 100 | 2 | Claus | BSR/Selectox | Yes | Yes | 0.233 | 483 | 175 | 450 |
| 21 | 555 | 3 | Claus | No | Yes | Yes | 0.592 | 6835 | 550 | 450 |
| 24 | 555 | 3 | Claus | BSR/MDEA | Yes | Yes | 0.009 | 121 | 100 | 450 |
| 27 | 555 | 3 | Claus | BSRP | No* | No | 0.007 | 51 | - | 100 |
| 30 | 555 | 3 | Claus | Sulfreen | Yes | Yes | 0.238 | 2734 | 350 | 450 |
| 33 | 555 | 2 | Claus | BSR/Selectox | Yes | Yes | 0.233 | 2678 | 350 | 450 |

---

*Combustor only

Table 1-4 (Table E-4)

### Onshore Sour Natural Gas Sulfur Recovery Study

### End-of-Run Sulfur Emission as $SO_2$ and Stack Height

### Acid Gas $H_2S/CO_2$ Ratio of 12.5/87.5

| | | SULFUR RECOVERY UNIT | | | THERMAL OXIDIZER | | SULFUR AS $SO_2$ | | STACK | |
|---|---|---|---|---|---|---|---|---|---|---|
| CASE NO. | SULFUR INPUT (LT/D) | NO. STAGES | TYPE | TAIL GAS UNIT | INCLUDED | WASTE HEAT RECOVERY | VOL % | LB/HR | HEIGHT (FT) | TEMP (°F) |
| 4A | 5 | 2 | Recycle Selectox | No | Yes | No | 0.654 | 101 | 100 | 1200 |
| 6A | 10 | 2 | Claus | No | Yes | No | 0.633 | 194 | 125 | 1200 |
| 6C | 10 | 2 | Recycle Selectox | No | Yes | No | 0.654 | 201 | 125 | 1200 |
| 8A | 10 | 3 | Claus | No | Yes | No | 0.483 | 148 | 100 | 1200 |
| 8C | 10 | 3 | Recycle Selectox | No | Yes | No | 0.424 | 130 | 100 | 1200 |
| 10A | 100 | 3 | Claus | No | Yes | Yes | 0.483 | 1480 | 300 | 450 |
| 10C | 100 | 3 | Recycle Selectox | No | Yes | Yes | 0.424 | 1300 | 275 | 450 |

## SECTION 2 (SECTION E-2)

## SUMMARY OF RESULTS

### EMISSIONS AND INVESTMENT COSTS

The emissions for the various cases are provided as outlined below.

The sulfur emissions as $SO_2$ vol % and lb/hr along with stack heights are shown in Tables 1-2, 1-3, and 1-4. They are also given as ppmv and again as lb/hr compared with total investment costs in Tables 5-1, 5-2, and 5-3.

The compositions of the exit gases in lb mols/hr are included in Tables 2-1, 2-2, and 2-3.

Separate investment costs for each sulfur recovery process, tail gas process, and thermal oxidizer, waste heat boiler, and stack are provided in Tables 5-4, 5-5, and 5-6.

# Table 2-1 (Table E-5)

## Onshore Sour Natural Gas Sulfur Recovery Study

## Composition of Exit Gases (lb mols/hr) End-of-Run

## Acid Gas $H_2S/CO_2$ Ratios of 80/20 and 50/50

| CASE NO. | SULFUR INPUT (LT/D) | SULFUR RECOVERY UNIT |  | TAIL GAS UNIT | THERMAL OXIDIZER |  | TAIL GAS COMPOSITION (lb mols/hr) |  |  |  |  |  |  |  |  |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
|  |  | NO. STAGES | TYPE |  | INCLUDED | WASTE HEAT RECOVERY | $SO_2$ | $O_2$ | $N_2$ | $CO_2$ | $H_2$ | CO | COS | $H_2O$ | TOTAL |
| Acid Gas $H_2S/CO_2$ Ratio - 80/20 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 19 | 1,000 | 3 |  | No | Yes | Yes | 117.30 | 143.60 | 7890.70 | 939.40 | - | - | - | 143.60 | 9234.60 |
| Acid Gas $H_2S/CO_2$ Ratio - 50/50 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 1 | 5 | - | No | No | Yes | No | 14.56 | 32.72 | 206.47 | 14.56 | - | - | - | 24.98 | 293.29 |
| 3 | 5 | 2 | Recycle Selectox | No | Yes | No | 1.10 | 1.08 | 27.32 | 32.97 | - | - | - | 22.60 | 85.07 |
| 5 | 10 | 2 | Claus | No | Yes | No | 2.15 | 2.00 | 90.43 | 32.16 | - | - | - | 42.78 | 169.52 |
| 7 | 10 | 3 | Claus | No | Yes | No | 1.38 | 1.98 | 90.00 | 32.37 | - | - | - | 43.18 | 168.91 |
| 9 | 100 | 3 | Claus | No | Yes | Yes | 13.81 | 19.84 | 900.10 | 323.70 | - | - | - | 431.80 | 1689.25 |
| 11 | 100 | 3 | Claus | BSR/MDEA | Yes | Yes | 0.18 | 21.85 | 1134.00 | 365.50 | - | - | - | 176.90 | 1698.43 |
| 13 | 100 | 3 | Claus | BSRP | No* | No | - | - | 626.90 | 308.20 | 11.03 | 0.62 | 0.04 | 65.40 | 1012.19 |
| 15 | 100 | 3 | Claus | Sulfreen | Yes | Yes | 5.52 | 19.31 | 890.90 | 325.50 | - | - | - | 434.50 | 1675.73 |
| 17 | 100 | 2 | Claus | BSR/Selectox | Yes | Yes | 5.29 | 17.83 | 989.40 | 338.70 | - | - | - | 158.60 | 1509.82 |
| 20 | 1,000 | 3 | Claus | No | Yes | Yes | 138.10 | 198.40 | 9001.00 | 3237.00 | - | - | - | 4318.00 | 16892.50 |
| 23 | 1,000 | 3 | Claus | BSR/MDEA | Yes | Yes | 1.79 | 218.50 | 11,340.00 | 3655.00 | - | - | - | 1769.00 | 16984.29 |
| 26 | 1,000 | 3 | Claus | BSRP | No* | No | - | - | 6269.00 | 3082.00 | 110.30 | 6.20 | 0.36 | 653.80 | 10121.66 |
| 29 | 1,000 | 3 | Claus | Sulfreen | Yes | Yes | 55.24 | 193.10 | 8909.00 | 3255.00 | - | - | - | 4345.00 | 16757.34 |
| 32 | 1,000 | 2 | Claus | BSR/Selectox | Yes | Yes | 52.90 | 178.30 | 9894.00 | 3387.00 | - | - | - | 1586.00 | 15098.20 |

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*Combustor only

## Table 2-2 (Table E-6)

## Onshore Sour Natural Gas Sulfur Recovery Study

## Composition of Exit Gases (lb mols/hr) End-of-Run

## Acid Gas $H_2S/CO_2$ Ratio of 20/80

| CASE NO. | SULFUR INPUT (LT/D) | SULFUR RECOVERY UNIT | | TAIL GAS UNIT | THERMAL OXIDIZER | | TAIL GAS COMPOSITION (lb mols/hr) | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | NO. STAGES | TYPE | | INCLUDED | WASTE HEAT RECOVERY | $SO_2$ | $O_2$ | $N_2$ | $CO_2$ | $H_2$ | CO | COS | $H_2O$ | TOTAL |
| 2 | 5 | - | No | No | Yes | No | 14.56 | 18.58 | 155.00 | 58.24 | - | - | - | 26.46 | 272.84 |
| 4 | 5 | 2 | Recycle Selectox | No | Yes | No | 1.40 | 2.19 | 68.04 | 62.22 | - | - | - | 30.70 | 164.55 |
| 6 | 10 | 2 | Claus | No | Yes | No | 2.87 | 4.29 | 134.70 | 124.00 | - | - | - | 59.57 | 325.43 |
| 6B | 10 | 2 | Recycle Selectox | No | Yes | No | 2.80 | 4.38 | 136.10 | 124.40 | - | - | - | 61.40 | 329.08 |
| 8 | 10 | 3 | Claus | No | Yes | No | 1.92 | 4.26 | 134.20 | 124.30 | - | - | - | 60.08 | 324.76 |
| 8B | 10 | 3 | Recycle Selectox | No | Yes | No | 1.61 | 4.34 | 136.30 | 124.60 | - | - | - | 60.81 | 327.66 |
| 10 | 100 | 3 | Claus | No | Yes | Yes | 19.21 | 42.64 | 1342.0 | 1243.00 | - | - | - | 600.80 | 3247.65 |
| 10B | 100 | 3 | Recycle Selectox | No | Yes | Yes | 16.14 | 43.41 | 1363.00 | 1246.00 | - | - | - | 608.10 | 3276.65 |
| 12 | 100 | 3 | Claus | BSR/MDEA | Yes | Yes | 0.34 | 51.62 | 1896.00 | 1366.00 | - | - | - | 400.30 | 3714.26 |
| 14 | 100 | 3 | Claus | BSRP | No* | No | - | - | 767.90 | 1201.00 | 12.87 | 2.14 | 0.14 | 137.00 | 2121.05 |
| 16 | 100 | 3 | Claus | Sulfreen | Yes | Yes | 7.69 | 41.96 | 1330.00 | 1245.00 | - | - | - | 603.80 | 3228.45 |
| 18 | 100 | 2 | Claus | BSR/Selectox | Yes | Yes | 7.54 | 4.35 | 1598.00 | 1280.00 | - | - | - | 347.80 | 3237.69 |
| 21 | 555 | 3 | Claus | No | Yes | Yes | 106.70 | 236.70 | 7448.00 | 6899.00 | - | - | - | 3334.00 | 18024.40 |
| 24 | 555 | 3 | Claus | BSR/MDEA | Yes | Yes | 1.89 | 286.50 | 10520.00 | 7580.00 | - | - | - | 2222.00 | 20610.39 |
| 27 | 555 | 3 | Claus | BSRP | No* | No | - | - | 4262.00 | 6667.00 | 71.43 | 11.86 | 0.80 | 760.40 | 11773.49 |
| 30 | 555 | 3 | Claus | Sulfreen | Yes | Yes | 42.68 | 232.90 | 7379.00 | 6909.00 | - | - | - | 3351.00 | 17914.58 |
| 33 | 555 | 2 | Claus | BSR/Selectox | Yes | Yes | 41.85 | 24.14 | 8869.00 | 7104.00 | - | - | - | 1930.00 | 17968.99 |

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*Combuster only

## Table 2-3 (Table E-7)

## Onshore Sour Natural Gas Sulfur Recovery Study

## Composition of Exit Gases (lb mols/hr) End-of-Run

## Acid Gas $H_2S/CO_2$ Ratio of 12.5/87.5

| | | SULFUR RECOVERY UNIT | | | THERMAL OXIDIZER | | TAIL GAS COMPOSITION (lb mols/hr) | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| CASE NO. | SULFUR INPUT (LT/D) | NO. STAGES | TYPE | TAIL GAS UNIT | INCLUDED | WASTE HEAT RECOVERY | $SO_2$ | $O_2$ | $N_2$ | $CO_2$ | $H_2$ | CO | COS | $H_2O$ | TOTAL |
| 4A | 5 | 2 | Recycle Selectox | No | Yes | No | 1.57 | 3.25 | 88.41 | 108.00 | - | - | - | 38.80 | 240.03 |
| 6A | 10 | 2 | Claus | No | Yes | No | 3.03 | 6.49 | 177.00 | 215.80 | - | - | - | 76.29 | 478.61 |
| 6C | 10 | 2 | Recycle Selectox | No | Yes | No | 3.14 | 6.50 | 176.80 | 215.90 | - | - | - | 77.60 | 479.94 |
| 8A | 10 | 3 | Claus | No | Yes | No | 2.31 | 6.47 | 176.60 | 216.00 | - | - | - | 76.68 | 478.06 |
| 8C | 10 | 3 | Recycle Selectox | No | Yes | No | 2.03 | 6.50 | 177.00 | 216.20 | - | - | - | 76.85 | 478.58 |
| 10A | 100 | 3 | Claus | No | Yes | Yes | 23.11 | 64.71 | 1766.00 | 2160.00 | - | - | - | 766.80 | 4780.62 |
| 10C | 100 | 3 | Recycle Selectox | No | Yes | Yes | 20.30 | 65.04 | 1770.00 | 2162.00 | - | - | - | 768.50 | 4785.84 |

## SECTION 3 (SECTION E-3)

## SULFUR RECOVERY PROCESSES

### RECYCLE SELECTOX PROCESS DESCRIPTION

The Selectox catalyst enables $H_2S$ to be oxidized to sulfur with air at low temperature, eliminating the need for high-temperature combustion as in the Claus Sulfur Recovery Process.

When supplied with the proper amount of air, the oxidation of one-third of the $H_2S$ to $SO_2$ and reaction with the remaining two-thirds of $H_2S$ occur simultaneously in the presence of the Selectox catalyst to form elemental sulfur:

$$H_2S + 3/2\ O_2 \longrightarrow SO_2 + H_2O$$

$$2H_2S + SO_2 \longrightarrow 3S + 2H_2O$$

The effective overall reaction is therefore:

$$3H_2S + 3/2\ O_2 \longrightarrow 3S + 3H_2O$$

The process is called Selectox when the acid gas feed strength is 5% $H_2S$ or less. The exothermic reaction with 5% gas results in a reasonable maximum reactor outlet temperature. After the Selectox reactor, the gas is cooled in a sulfur condenser, which produces steam and condenses sulfur.

When the acid gas feed contains more than 5% $H_2S$, some of the cooled lean gas from the sulfur condenser is reheated and recycled to the Selectox reactor inlet to maintain approximately 5% $H_2S$ at the reactor inlet. This process is called Recycle Selectox.

Gas from the sulfur condenser proceeds through one or two Claus stages, each with a reheater, converter, and sulfur condenser.

The Recycle Selectox 2-Stage Process, shown in Figure 3-1, has one Selectox and one Claus stage. The Recycle Selectox 3-Stage Process, illustrated in Figure 3-2, is identical but has a second Claus stage added.

The system for the Selectox and the Claus reactors is the same: heating gas to the desired inlet temperature, reaction in a converter, and cooling the gas and condensing sulfur in a condenser. The condensers produce low-pressure steam.

The first industrial plant using Recycle Selectox is in an advanced state of construction and is expected to start operation about November 1981. Selectox catalyst has been proven in 3-1/2 years of operation in a relevant industrial plant.

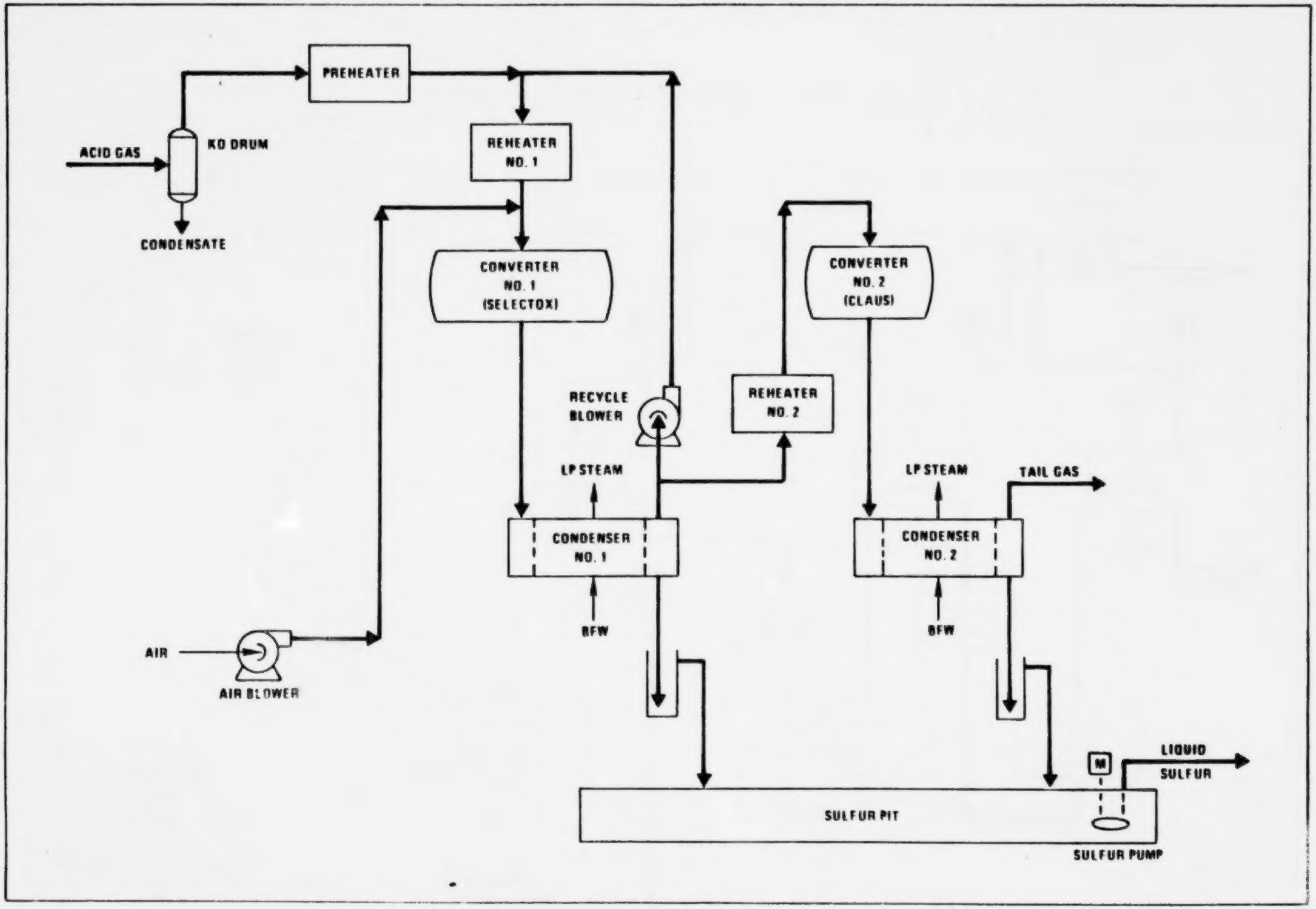
ACID GAS
KO DRUM
CONDENSATE
PREHEATER
REHEATER NO. 1
CONVERTER NO. 1 (SELECTOX)
AIR
AIR BLOWER
RECYCLE BLOWER
LP STEAM
CONDENSER NO. 1
BFW
REHEATER NO. 2
CONVERTER NO. 2 (CLAUS)
LP STEAM
TAIL GAS
CONDENSER NO. 2
BFW
M
LIQUID SULFUR
SULFUR PIT
SULFUR PUMP

Figure 3-1. Recycle Selectox 2-Stage Process
(Figure E-1)

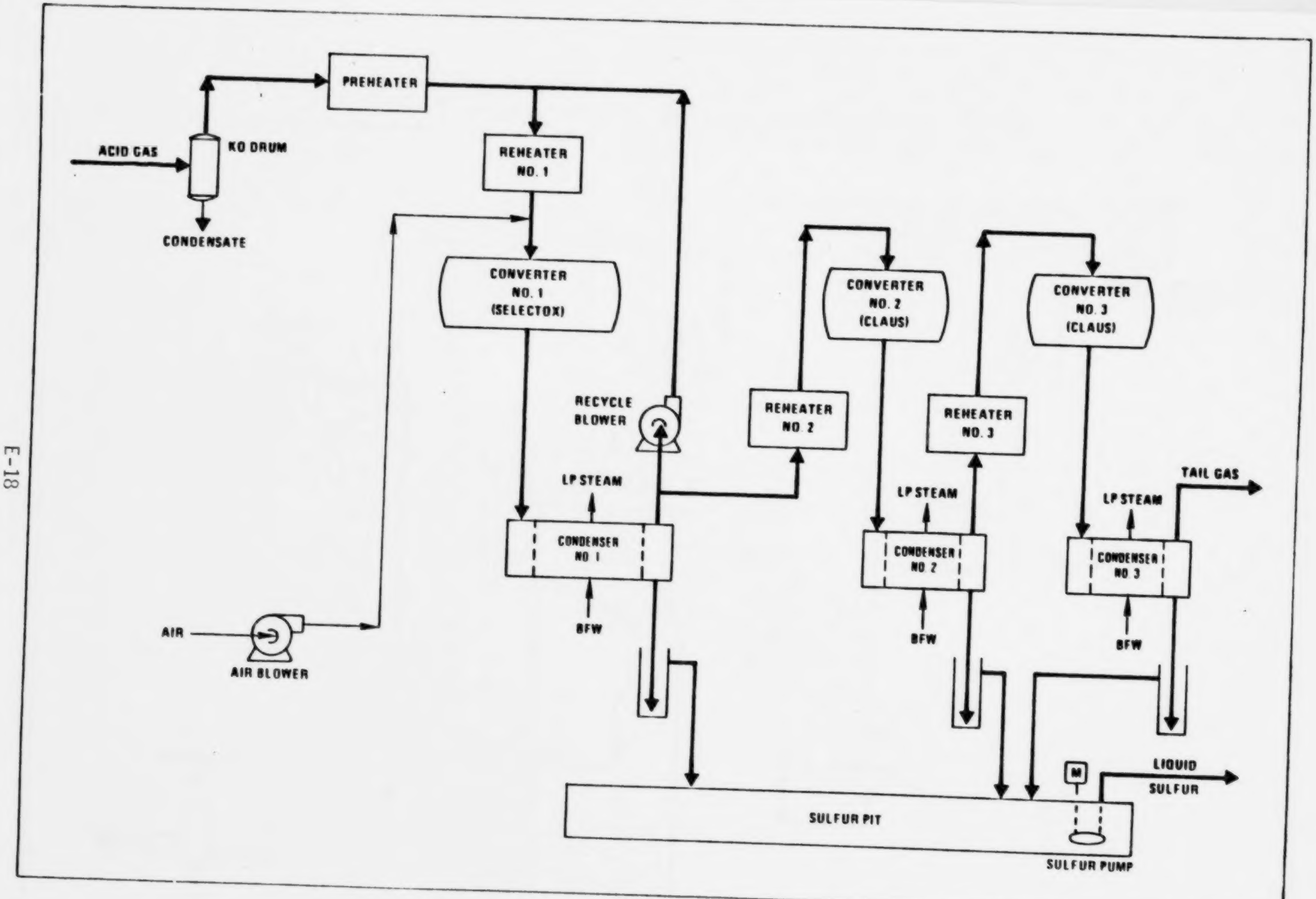
ACID GAS
KO DRUM
CONDENSATE
PREHEATER
REHEATER NO. 1
CONVERTER NO. 1 (SELECTOX)
AIR
AIR BLOWER
RECYCLE BLOWER
LP STEAM
CONDENSER NO 1
BFW
REHEATER NO. 2
CONVERTER NO. 2 (CLAUS)
LP STEAM
CONDENSER NO. 2
BFW
REHEATER NO. 3
CONVERTER NO. 3 (CLAUS)
LP STEAM
CONDENSER NO. 3
BFW
TAIL GAS
M
LIQUID SULFUR
SULFUR PIT
SULFUR PUMP

Figure 3-2. Recycle Selectox 3-Stage Process (Figure E-2)

## CLAUS PROCESS DESCRIPTION

The Claus reaction consists of combining 2 mols of $H_2S$ with 1 mol of $SO_2$ to form elemental sulfur:

$$2H_2S + SO_2 \longrightarrow 3S + 2H_2O \qquad (1)$$

Normally, for gases rich in $H_2S$ (50% to 80%), in the so-called Claus Process the acid gas is burned with the proper amount of air to oxidize one-third of the $H_2S$ to make the required amount of $SO_2$ to combine the remaining two-thirds to create the following reaction:

$$H_2S + 3/2\ O_2 \longrightarrow SO_2 + H_2O \qquad (2)$$

This makes a high enough temperature to trigger the reaction thermally. However, when acid gas is weak in $H_2S$, such as in the cases with $H_2S/CO_2$ratios of 20/80 and 12.5/87.5, the $CO_2$ dilution makes the temperature in the reaction furnace too low. A sulfur-burning modification is therefore used to make the required $SO_2$ by burning recovered sulfur in the reaction furnace:

$$S + O_2 \longrightarrow SO_2 \qquad (3)$$

The preheated acid gas containing all of the $H_2S$ goes directly to the first converter, where it reacts with the $SO_2$ made in the reaction furnace. The first converter is followed by one or two additional converters.

The Claus 2-stage process is illustrated in Figure 3-3. The acid gas flows through a knockout drum to trap entrained liquid or slugs. The gas enters the reaction furnace along with air, which is ratio-controlled to burn one-third of the $H_2S$ to form $SO_2$. The gases are cooled in the reaction cooler, which produces high-pressure steam, and are further cooled in Condenser 1, which produces low-pressure steam. Sulfur is condensed and flows through a seal into the sulfur pit. The gas is heated in Reheater 1, flows through Converter 1 and through Condenser 2. This constitutes the first catalytic stage. The second catalytic stage is identical. The sulfur collected from all three condensers flows into the sulfur pit.

The Claus 3-stage process, shown in Figure 3-4, is identical to the 2-stage process in Figure 3-3 plus the addition of a reheater, a converter and a condenser.

The Claus sulfur-burning 2-stage process, shown in Figure 3-5, is the same as the 2-stage process in Figure 3-3, except that sulfur instead of acid gas is burned in the reaction furnace, and the acid gas is preheated and goes to Converter 1.

The Claus sulfur-burning 3-stage process, shown in Figure 3-6, is identical to the 2-stage process in Figure 3-5 plus the addition of a reheater, a converter and a condenser.

The acid gas preheaters are usually heated by steam. The reheaters can be either auxiliary burners, which heat the process gases with either acid gas or fuel gas, or steam-heated shell-and-tube units.

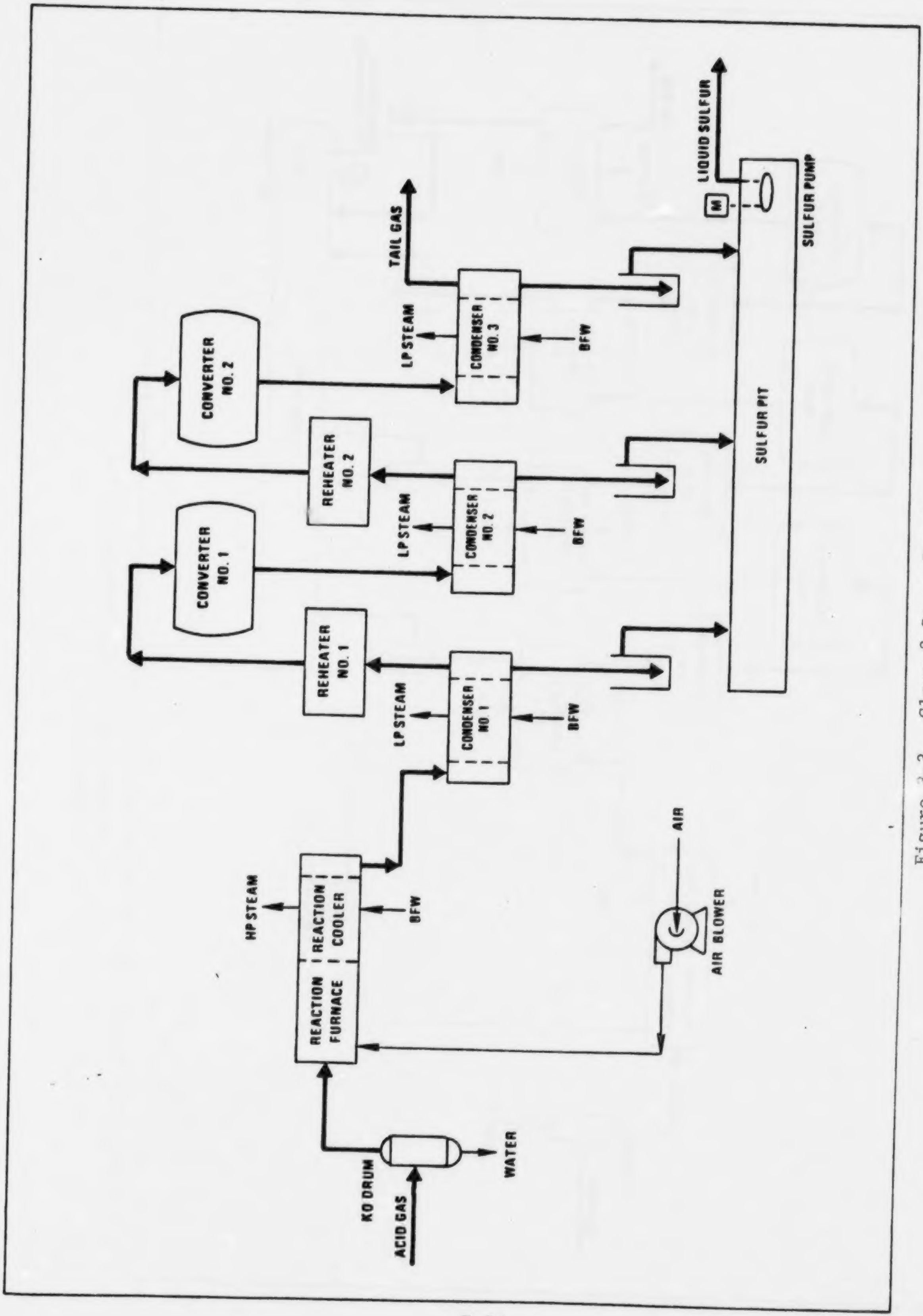
ACID GAS
KO DRUM
WATER
REACTION FURNACE
REACTION COOLER
HP STEAM
BFW
AIR BLOWER
AIR
CONDENSER NO. 1
LP STEAM
BFW
REHEATER NO. 1
CONVERTER NO. 1
CONDENSER NO. 2
LP STEAM
BFW
REHEATER NO. 2
CONVERTER NO. 2
CONDENSER NO. 3
LP STEAM
BFW
TAIL GAS
SULFUR PIT
M
SULFUR PUMP
LIQUID SULFUR

Figure 3-3. Claus 2-Stage Process
(Figure E-3)

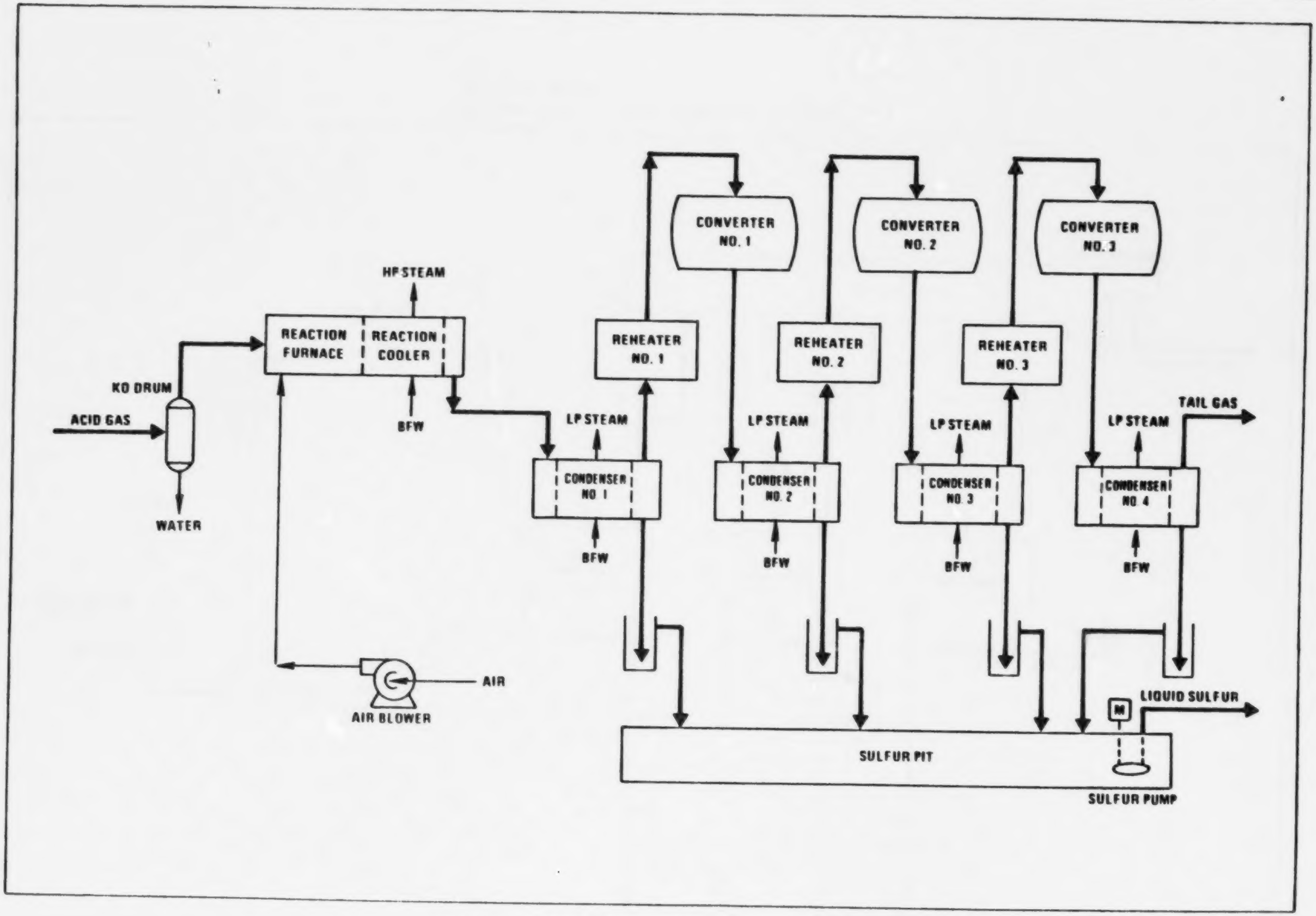
ACID GAS
KO DRUM
WATER
REACTION FURNACE
REACTION COOLER
HP STEAM
BFW
AIR BLOWER
AIR
LP STEAM
CONDENSER NO. 1
BFW
REHEATER NO. 1
CONVERTER NO. 1
LP STEAM
CONDENSER NO. 2
BFW
REHEATER NO. 2
CONVERTER NO. 2
LP STEAM
CONDENSER NO. 3
BFW
REHEATER NO. 3
CONVERTER NO. 3
LP STEAM
CONDENSER NO. 4
BFW
TAIL GAS
M
LIQUID SULFUR
SULFUR PIT
SULFUR PUMP

Figure 3-4. Claus 3-Stage Process
(Figure E-4)

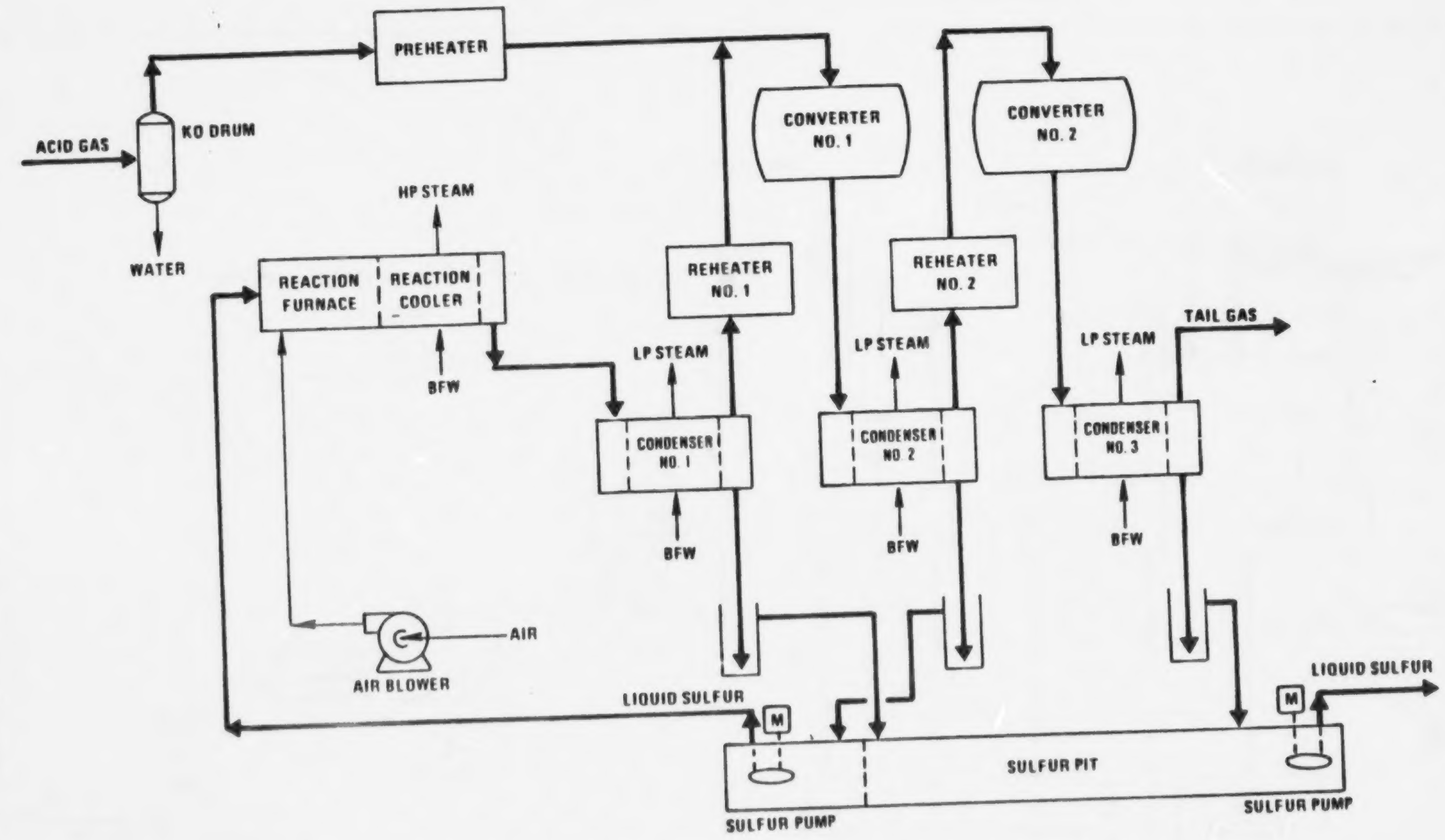
ACID GAS
KO DRUM
WATER
PREHEATER
HP STEAM
REACTION FURNACE
REACTION COOLER
BFW
CONVERTER NO. 1
CONVERTER NO. 2
REHEATER NO. 1
REHEATER NO. 2
LP STEAM
CONDENSER NO. 1
LP STEAM
CONDENSER NO. 2
LP STEAM
CONDENSER NO. 3
TAIL GAS
BFW
BFW
BFW
AIR
AIR BLOWER
LIQUID SULFUR
M
SULFUR PIT
M
LIQUID SULFUR
SULFUR PUMP
SULFUR PUMP

Figure 3-5. Claus Sulfur-Burning 2-Stage Process
(Figure E-5)

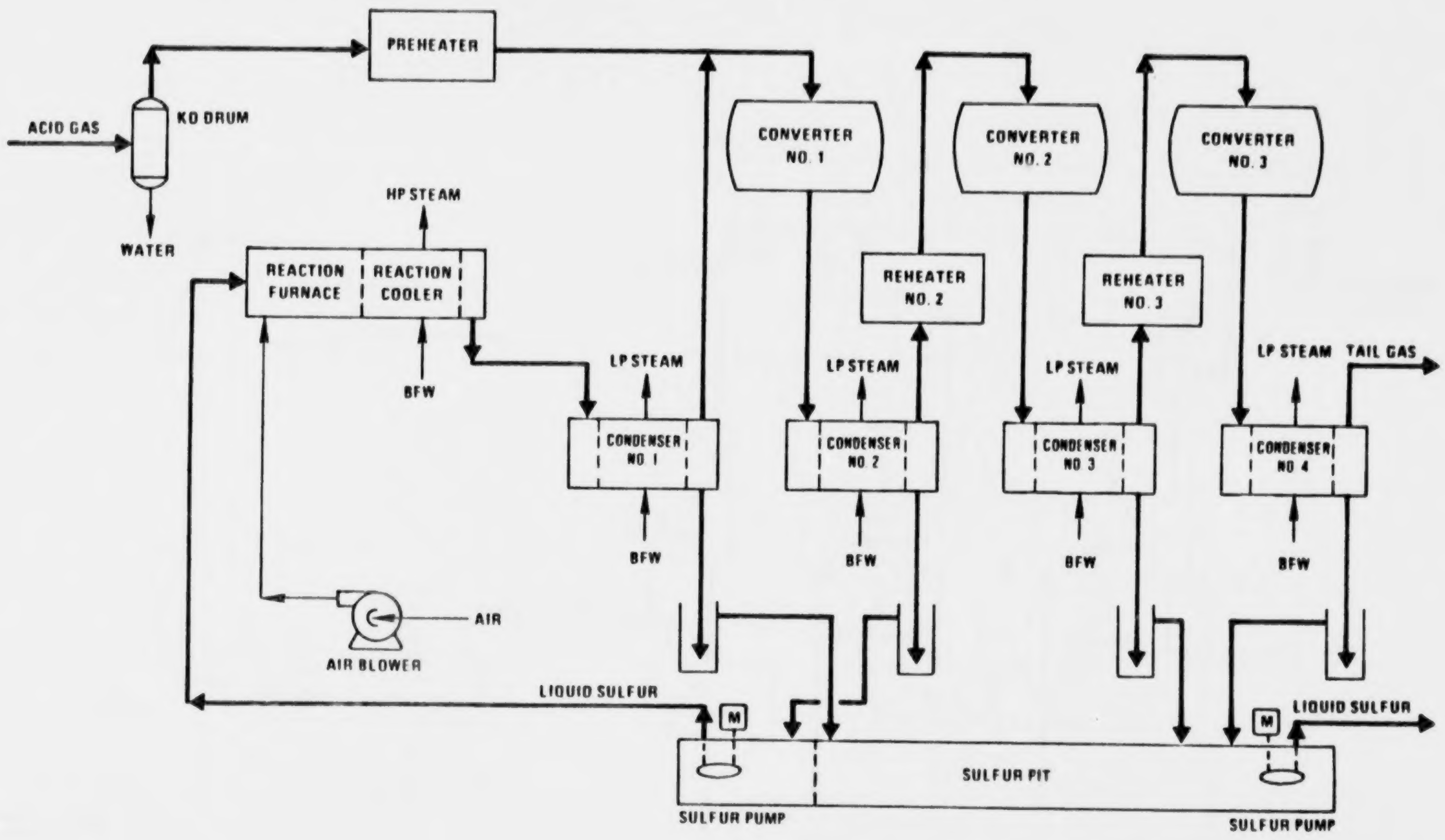
ACID GAS
KO DRUM
WATER
PREHEATER
HP STEAM
REACTION FURNACE
REACTION COOLER
BFW
AIR BLOWER
AIR
CONVERTER NO. 1
CONVERTER NO. 2
CONVERTER NO. 3
REHEATER NO. 2
REHEATER NO. 3
LP STEAM
LP STEAM
LP STEAM
LP STEAM
TAIL GAS
CONDENSER NO. 1
CONDENSER NO. 2
CONDENSER NO. 3
CONDENSER NO. 4
BFW
BFW
BFW
BFW
LIQUID SULFUR
M
M
LIQUID SULFUR
SULFUR PIT
SULFUR PUMP
SULFUR PUMP

Figure 3-6. Claus Sulfur-Burning 3-Stage Process
(Figure E-6)

## SECTION 4 (SECTION E-4)

## TAIL GAS PROCESSES

### THERMAL OXIDIZERS, WASTE HEAT BOILERS, AND STACKS

The two types of arrangements for burning the combustibles in tail gases to convert all sulfur compounds and sulfur vapor into $SO_2$ are shown in Figure 4-1. The combustion takes place with 25% excess air and the addition of the required amount of fuel gas to maintain a temperature of 1,200°F. This temperature is considered necessary to ensure that the stack gases will contain less than 10 ppm of $H_2S$.

For Cases 1 and 2, which have no sulfur recovery, the acid gas is thermally oxidized directly with no fuel gas required. It is necessary to use more than 25% excess air to limit the stack temperatures to 1,500°F.

For small plants, which have no waste heat recovery such as in Cases 1 through 8C, the thermal oxidizer is in the base of the stack. The stack provides more than sufficient residence time for completion of reactions. This is shown on the left side of Figure 4-1.

For all other cases, except BSRP Cases 13, 14, 26, and 27, which only require an emergency combustor, the scheme shown on the right side of Figure 4-1 is used. The thermal oxidizer must have sufficient residence time to ensure achieving complete combustion before entering the waste heat boiler where the gas is being cooled. Because the pressure drop through the waste heat boiler imposes a slight pressure on the thermal oxidizer, an air blower is required.

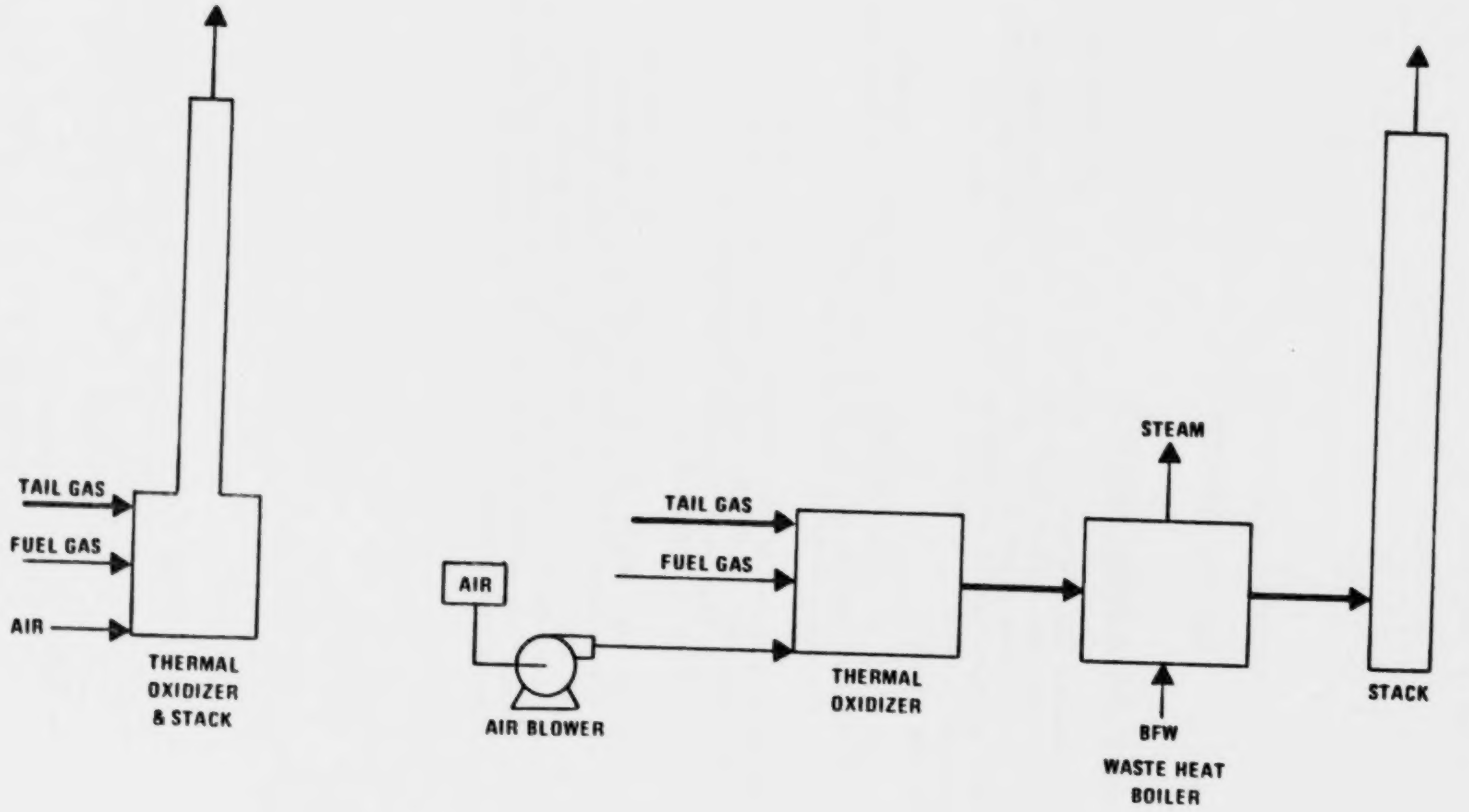
TAIL GAS
FUEL GAS
AIR
THERMAL OXIDIZER & STACK
AIR
AIR BLOWER
TAIL GAS
FUEL GAS
THERMAL OXIDIZER
STEAM
BFW
WASTE HEAT BOILER
STACK

Figure 4-1. Thermal Oxidizers, Waste Heat Boilers, and Stacks
(Figure E-7)

## BSR/MDEA PROCESS DESCRIPTION

The flow diagram for the BSR/MDEA Process is provided in Figure 4-2. The BSR or hydrogenation section is identical to the corresponding section in the BSR Process, except that it is somewhat larger to handle the recycled gas.

In the BSR section the sulfur recovery unit tail gas is heated to reaction temperature in the reducing gas generator by mixing it with the products of combustion of fuel gas and air. Some hydrogen and carbon monoxide are formed to supplement the hydrogen in the tail gas. The gas enters the hydrogenation reactor, where all sulfur compounds ($SO_2$, $S_x$, COS, $CS_2$) are converted to $H_2S$, and is then cooled in the reactor effluent cooler, which produces steam. The gas is further cooled in the contact condenser by evaporating water in the lower section and condensing and cooling in the upper section.

The cooled gas enters the MDEA absorber, where essentially all of the $H_2S$ and some of the $CO_2$ are absorbed by cool MDEA solution, which enters at the top. The gas, containing a small amount of $H_2S$, goes to the thermal oxidizer, waste heat boiler and stack.

The rich MDEA from the bottom of the absorber is preheated by lean MDEA and fed to the MDEA regenerator, where the $H_2S$ and $CO_2$ are stripped from the MDEA solution. This acid gas is recycled to the Claus unit.

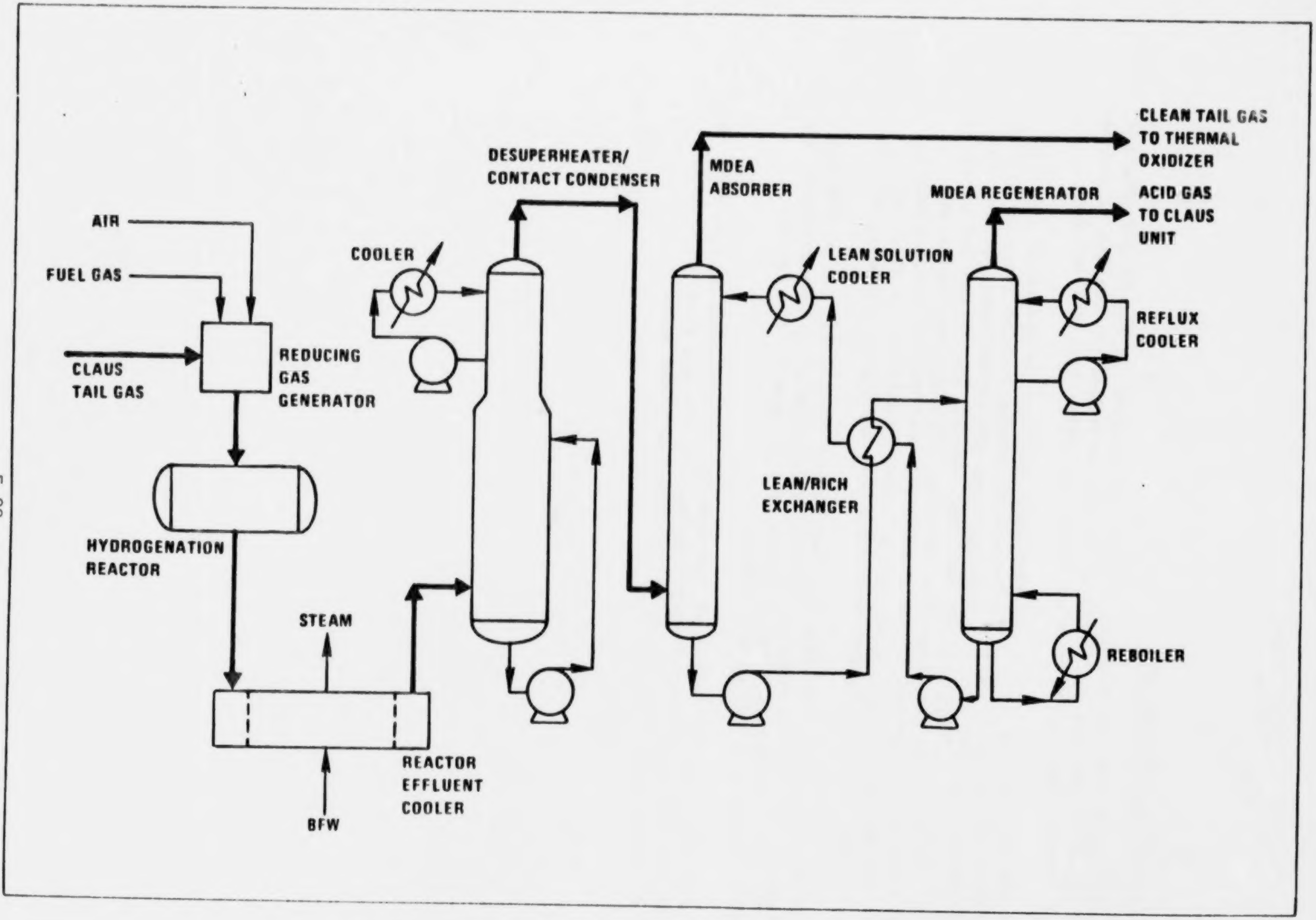
AIR
FUEL GAS
CLAUS TAIL GAS
REDUCING GAS GENERATOR
HYDROGENATION REACTOR
STEAM
BFW
REACTOR EFFLUENT COOLER
COOLER
DESUPERHEATER/ CONTACT CONDENSER
MDEA ABSORBER
LEAN SOLUTION COOLER
LEAN/RICH EXCHANGER
MDEA REGENERATOR
REFLUX COOLER
REBOILER
CLEAN TAIL GAS TO THERMAL OXIDIZER
ACID GAS TO CLAUS UNIT

Figure 4-2. BSR/MDEA Process
(Figure E-8)

## BEAVON SULFUR REMOVAL PROCESS (BSRP) DESCRIPTION

The flow diagram for the Beavon Sulfur Removal Process (BSRP) is shown in Figure 4-3. The process consists of the BSR or hydrogenation section and the Stretford section.

In the BSR section, the sulfur recovery unit tail gas is heated to reaction temperature in the reducing gas generator by mixing it with the products of combustion of fuel gas and air. Some hydrogen and carbon monoxide are formed to supplement the hydrogen in the tail gas. The gas enters the hydrogenation reactor, where essentially all sulfur compounds ($SO_2$, $S_x$, COS, $CS_2$) are converted to $H_2S$, and is then cooled in the reactor effluent cooler, which produces steam. The gas is further cooled in the contact condenser by evaporating water in the lower section and condensing and cooling in the upper section.

In the Stretford section the cooled hydrogenated gas is contacted in a venturi scrubber and an absorber, where the $H_2S$ is absorbed in an oxidizing alkaline solution. The $H_2S$ is converted to elemental sulfur by the oxidizing agents in the solution. The solution is regenerated by contacting with air in the oxidizer tank where sulfur is floated off as a slurry. After separating the sulfur from the chemicals by filtering or centrifuging, with water washing, it is reslurried with wash water and heated to melt the sulfur. The molten sulfur flows from the decanter to the sulfur pit. The chemicals are returned to the system and the wash water is discarded.

The offgas from the absorber typically contains less than 1 ppm of $H_2S$ and, therefore, does not need to be thermally oxidized. A combustor and short stack on top of the absorber are provided for rare occasions when the $H_2S$ content exceeds 10 ppm.

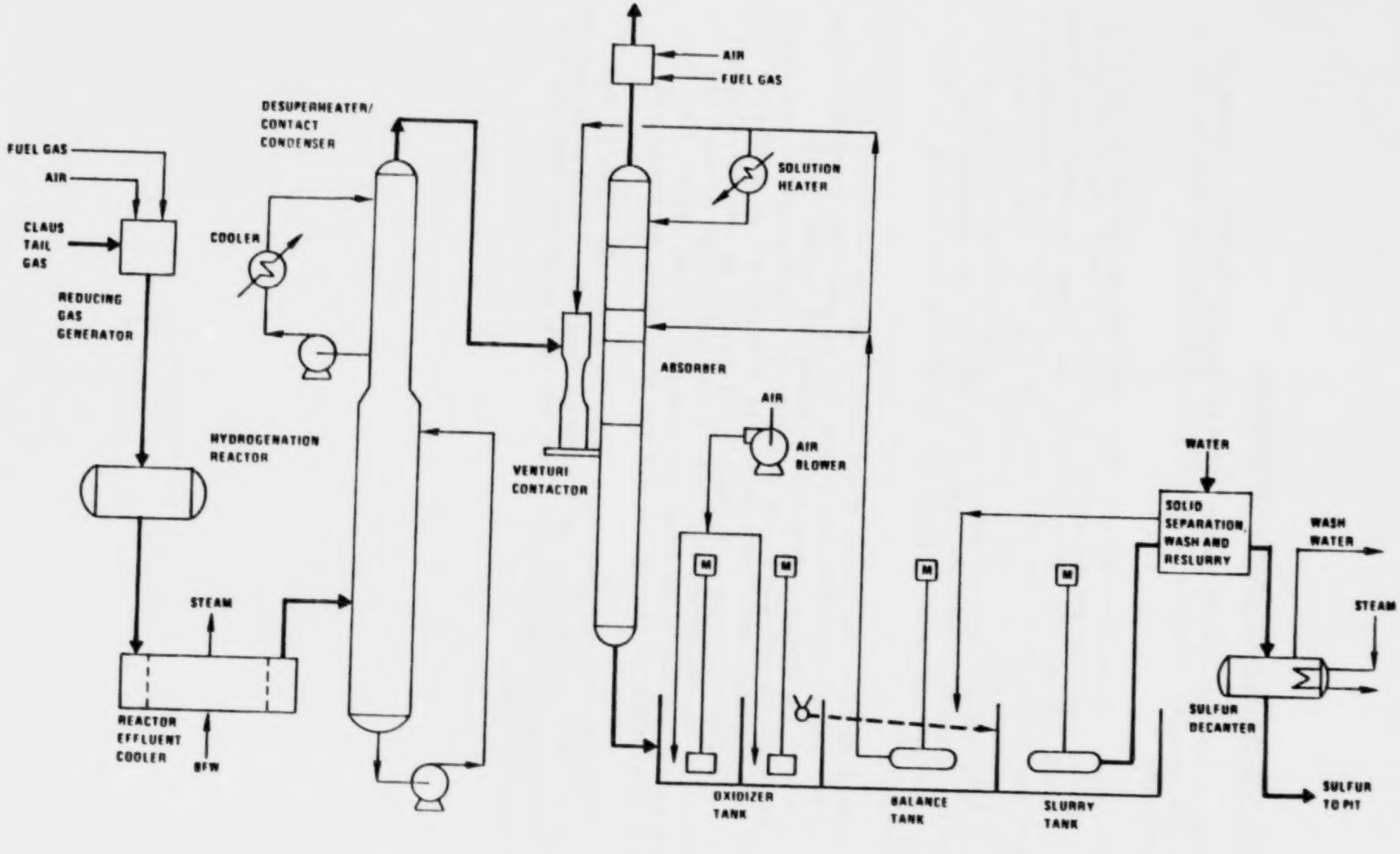
FUEL GAS
AIR
CLAUS TAIL GAS
REDUCING GAS GENERATOR
HYDROGENATION REACTOR
STEAM
REACTOR EFFLUENT COOLER
BFW
COOLER
DESUPERHEATER/ CONTACT CONDENSER
VENTURI CONTACTOR
AIR
FUEL GAS
SOLUTION HEATER
ABSORBER
AIR
AIR BLOWER
M
M
M
M
OXIDIZER TANK
BALANCE TANK
SLURRY TANK
WATER
SOLID SEPARATION, WASH AND RESLURRY
WASH WATER
STEAM
SULFUR DECANTER
SULFUR TO PIT

Figure 4-3. Beavon Sulfur Removal Process (BSRP)
(Figure E-9)

## SULFREEN PROCESS DESCRIPTION

The Sulfreen Process, shown in Figure 4-4, takes advantage of furthering the Claus reaction by operating below the sulfur dewpoint of the reaction gas mixture:

$$2H_2S + SO_2 \longrightarrow 3S + 2H_2O$$

Liquid sulfur is adsorbed on the catalyst which removes it from the gas, thereby allowing the reaction to move further to the right and obtaining a higher conversion than in the Claus process where the tail gas is above the sulfur dewpoint.

The catalyst is alumina, usually the same type used in the Claus process.

After a period of operation when the catalyst has adsorbed its limit of sulfur, the reactor enters a closed-cycle desorption operation, where recyle gas is heated and passed through the reactor to strip sulfur from the catalyst, then through a sulfur condenser. This condenser cools the gas and generates low-pressure steam. Sulfur is condensed and drained to the sulfur pit.

There are usually three reactors, two of which are in adsorption service while one is being regenerated.

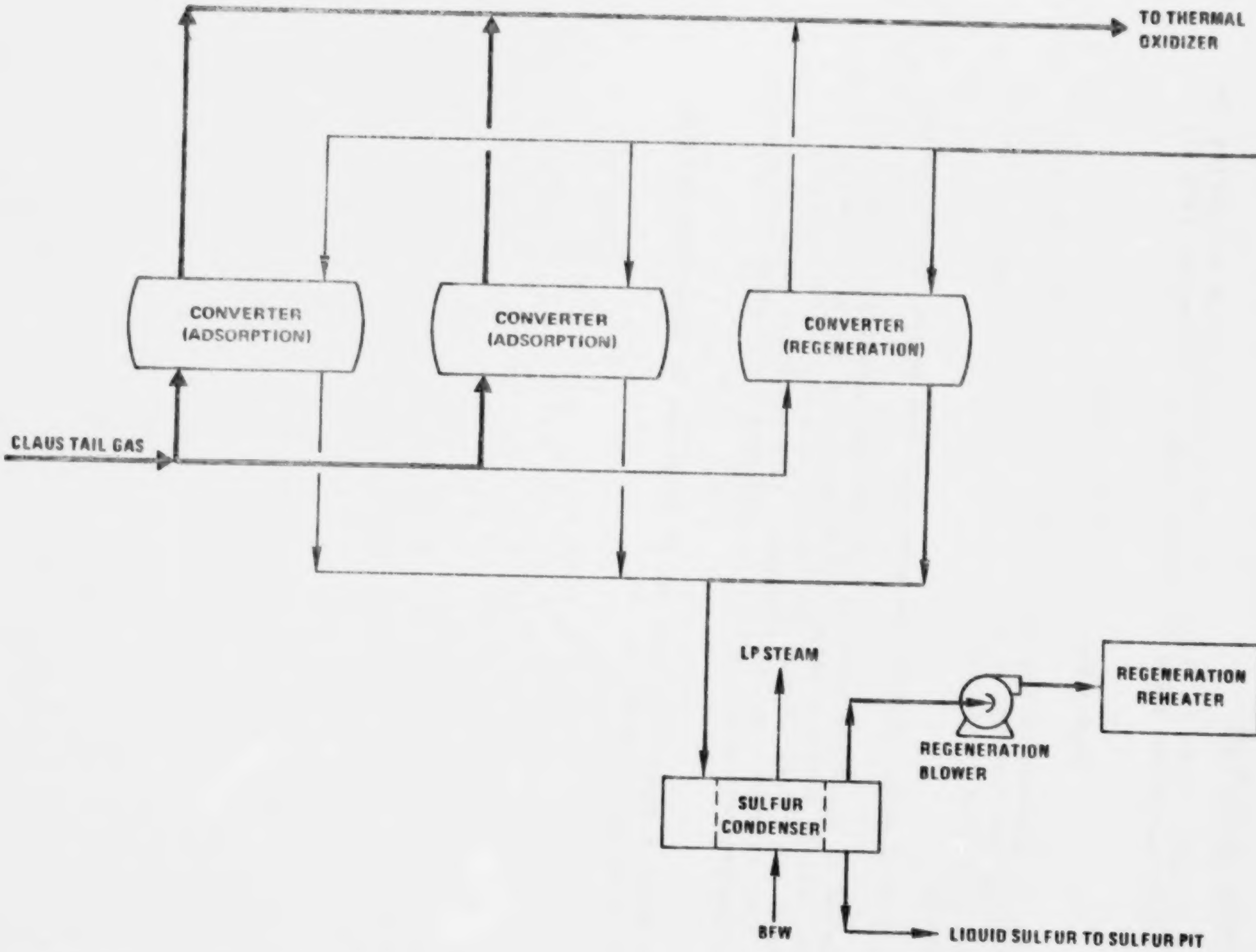
TO THERMAL OXIDIZER
CONVERTER (ADSORPTION)
CONVERTER (ADSORPTION)
CONVERTER (REGENERATION)
CLAUS TAIL GAS
LP STEAM
REGENERATION REHEATER
REGENERATION BLOWER
SULFUR CONDENSER
BFW
LIQUID SULFUR TO SULFUR PIT

Figure 4-4. Sulfreen Process
(Figure E-10)

## BSR/SELECTOX PROCESS DESCRIPTION

The flow diagram for the BSR/Selectox Process is provided in Figure 4-5. The process consists of a BSR or hydrogenation section and a Selectox section.

In the BSR section the sulfur recovery unit tail gas is heated to reaction temperature in the reducing gas generator by mixing it with the products of combustion of fuel gas and air. Some hydrogen and carbon monoxide are formed to supplement the hydrogen in the tail gas. The gas enters the hydrogenation reactor, where all sulfur compounds ($SO_2$, $S_x$, COS, $CS_2$) are converted to $H_2S$, and is then cooled in the reactor effluent, which produces steam. The gas is further cooled in the condenser by circulating water countercurrent to the gas. The purpose of cooling is to increase the potential conversion by removing water, which is one of the products of reaction.

The gas from the condenser is reheated to a moderate temperature and combined with air, after which it enters the Selectox reactor where the $H_2S$ is directly oxidized to elemental sulfur:

$$H_2S + 1/2\ O_2 \longrightarrow S + H_2O$$

The gas is cooled by generating low-pressure steam, and the condensed sulfur flows to the sulfur pit. The tail gas from the sulfur condenser goes to a thermal oxidizer, waste heat boiler and stack.

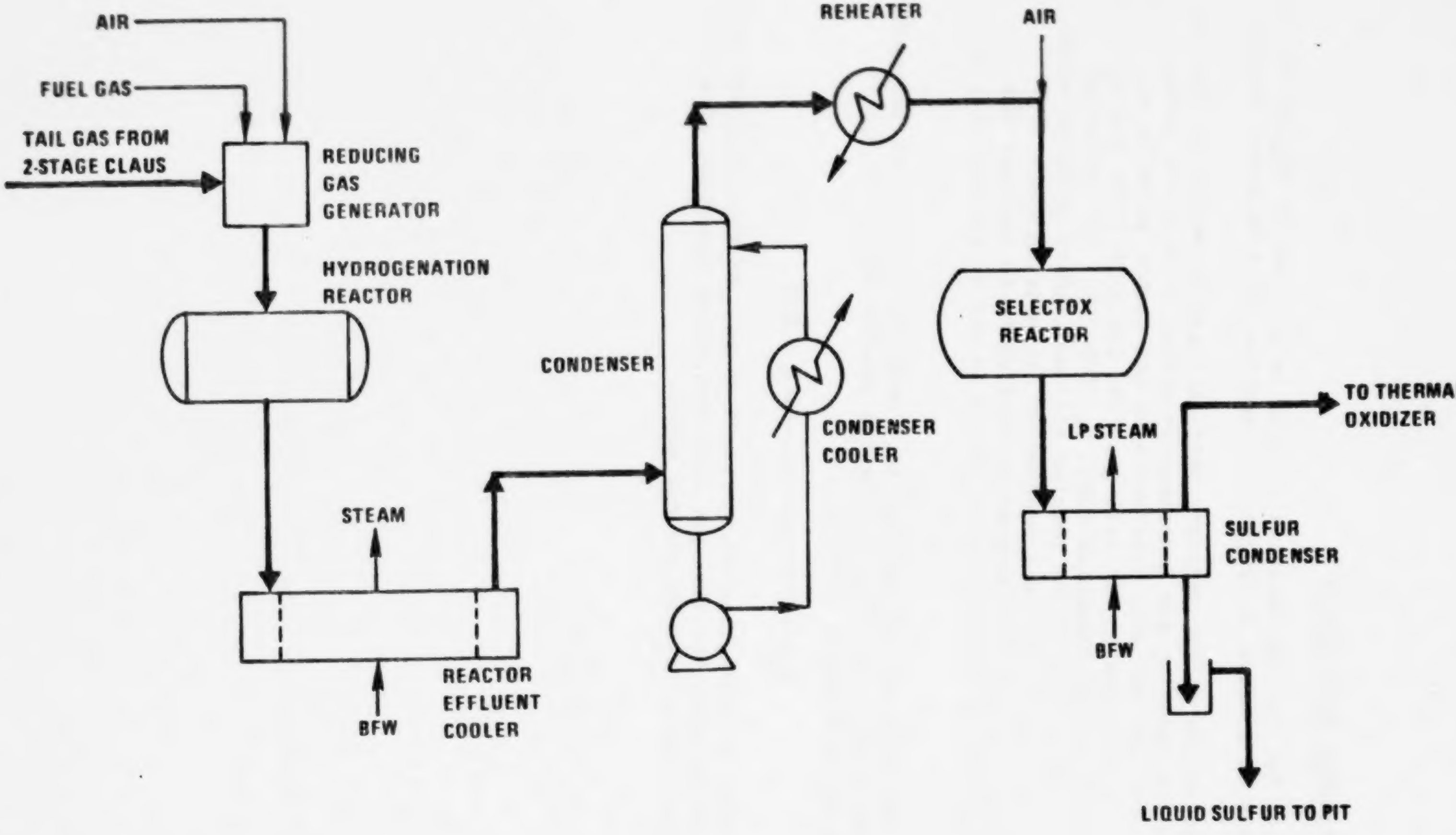
AIR
FUEL GAS
TAIL GAS FROM
2-STAGE CLAUS
REDUCING
GAS
GENERATOR
HYDROGENATION
REACTOR
STEAM
BFW
REACTOR
EFFLUENT
COOLER
CONDENSER
CONDENSER
COOLER
REHEATER
AIR
SELECTOX
REACTOR
LP STEAM
BFW
SULFUR
CONDENSER
TO THERMAL
OXIDIZER
LIQUID SULFUR TO PIT

Figure 4-5. BSR/SELECTOX Process
(Figure E-11)

## SECTION 5 (SECTION E-5)

## INVESTMENT COSTS

### INVESTMENT COSTS VS. SULFUR EMISSIONS

The total investment costs are compared with sulfur emissions for each case in Tables 5-1, 5-2, and 5-3. These investment costs are based on January 1981 Gulf Coast prices and include all design, engineering, purchasing of equipment and materials, and construction costs and contractors' fees, but do not include the initial charge of catalysts and chemicals. The investment costs were developed separately for each sulfur recovery process, tail gas process, and thermal oxidizer, waste heat boiler, and stack. These separate costs are provided in Tables 5-4, 5-5, and 5-6.

The costs for the initial charge of catalysts and chemicals are given in Tables 5-7, 5-8, and 5-9.

Table 5-1 (Table E-8)

Onshore Sour Natural Gas Sulfur Recovery Study

Investment Costs and Sulfur Emissions (End-of-Run)

Acid Gas $H_2S/CO_2$ Ratios of 80/20 and 50/50

| CASE NO. | SULFUR INPUT (LT/D) | SULFUR RECOVERY UNIT |  | TAIL GAS UNIT | THERMAL OXIDIZER |  | INVESTMENT COST (MM$) | SULFUR EMISSION AS $SO_2$ |  |
|---|---|---|---|---|---|---|---|---|---|
|  |  | NO. STAGES | TYPE |  | INCLUDED | WASTE HEAT RECOVERY |  | VOL. PPM | LB/HR |
| **Acid Gas $H_2S/CO_2$ Ratio - 80/20** |  |  |  |  |  |  |  |  |  |
| 19 | 1,000 | 3 |  | No | Yes | Yes | 22.30 | 12,700 | 7,514 |
| **Acid Gas $H_2S/CO_2$ Ratio - 50/50** |  |  |  |  |  |  |  |  |  |
| 1 | 5 | - | No | No | Yes | No | 1.28 | 49,640 | 933 |
| 3 | 5 | 2 | Recycle Selectox | No | Yes | No | 2.43 | 12,930 | 70 |
| 5 | 10 | 2 | Claus | No | Yes | No | 2.95 | 12,680 | 138 |
| 7 | 10 | 3 | Claus | No | Yes | No | 3.32 | 8,170 | 88 |
| 9 | 100 | 3 | Claus | No | Yes | Yes | 6.47 | 8,170 | 884 |
| 11 | 100 | 3 | Claus | BSR/MDEA | Yes | Yes | 11.90 | 110 | 12 |
| 13 | 100 | 3 | Claus | BSRP | No* | No | 9.12 | 40 | 3 |
| 15 | 100 | 3 | Claus | Sulfreen | Yes | Yes | 8.26 | 3,290 | 354 |
| 17 | 100 | 2 | Claus | BSR/Selectox | Yes | Yes | 7.47 | 3,500 | 339 |
| 20 | 1,000 | 3 | Claus | No | Yes | Yes | 26.10 | 8,170 | 8,847 |
| 23 | 1,000 | 3 | Claus | BSR/MDEA | Yes | Yes | 47.17 | 110 | 115 |
| 26 | 1,000 | 3 | Claus | BSRP | No* | No | 29.99 | 40 | 23 |
| 29 | 1,000 | 3 | Claus | Sulfreen | Yes | Yes | 29.41 | 3,300 | 3,539 |
| 32 | 1,000 | 2 | Claus | BSR/Selectox | Yes | Yes | 28.93 | 3,500 | 3,389 |

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*Combustor only

Table 5-2 (Table E-9)

## Onshore Sour Natural Gas Sulfur Recovery Study

## Investment Costs and Sulfur Emissions (End-of-Run)

## Acid Gas $H_2S/CO_2$ Ratio of 20/80

| CASE NO. | SULFUR INPUT (LT/D) | SULFUR RECOVERY UNIT |  | TAIL GAS UNIT | THERMAL OXIDIZER |  | INVESTMENT COST (MM$) | SULFUR EMISSION AS $SO_2$ |  |
|---|---|---|---|---|---|---|---|---|---|
|  |  | NO. STAGES | TYPE |  | INCLUDED | WASTE HEAT RECOVERY |  | VOL. PPM | LB/HR |
| 2 | 5 | - | No | No | Yes | No | 1.28 | 53,400 | 933 |
| 4 | 5 | 2 | Recycle Selectox | No | Yes | No | 2.43 | 8,510 | 90 |
| 6 | 10 | 2 | Claus | No | Yes | No | 3.40 | 8,820 | 184 |
| 6B | 10 | 2 | Recycle Selectox | No | Yes | No | 2.67 | 8,510 | 179 |
| 8 | 10 | 3 | Claus | No | Yes | No | 3.74 | 5,910 | 123 |
| 8B | 10 | 3 | Recycle Selectox | No | Yes | No | 3.11 | 4,910 | 103 |
| 10 | 100 | 3 | Claus | No | Yes | Yes | 9.05 | 5,910 | 1,203 |
| 10B | 100 | 3 | Recycle Selectox | No | Yes | Yes | 8.46 | 4,910 | 1,034 |
| 12 | 100 | 3 | Claus | BSR/MDEA | Yes | Yes | 20.44 | 90 | 22 |
| 14 | 100 | 3 | Claus | BSRP | No* | No | 11.53 | 70 | 9 |
| 16 | 100 | 3 | Claus | Sulfreen | Yes | Yes | 10.88 | 2,380 | 493 |
| 18 | 100 | 2 | Claus | BSR/Selectox | Yes | Yes | 10.45 | 2,330 | 483 |
| 21 | 555 | 3 | Claus | No | Yes | Yes | 26.23 | 5,920 | 6,835 |
| 24 | 555 | 3 | Claus | BSR/MDEA | Yes | Yes | 66.35 | 90 | 121 |
| 27 | 555 | 3 | Claus | BSRP | No* | No | 29.42 | 70 | 51 |
| 30 | 555 | 3 | Claus | Sulfreen | Yes | Yes | 29.63 | 2,380 | 2,734 |
| 33 | 555 | 2 | Claus | BSR/Selectox | Yes | Yes | 29.89 | 2,330 | 2,678 |

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*Combustor only

Table 5-3 (Table E-10)

## Onshore Sour Natural Gas Sulfur Recovery Study

## Investment Costs and Sulfur Emissions (End-of-Run)

## Acid Gas $H_2S/CO_2$ Ratio of 12.5/87.5

| CASE NO. | SULFUR INPUT (LT/D) | SULFUR RECOVERY UNIT | | TAIL GAS UNIT | THERMAL OXIDIZER | | INVESTMENT COST (MM$) | SULFUR EMISSION AS $SO_2$ | |
|---|---|---|---|---|---|---|---|---|---|
| | | NO. STAGES | TYPE | | INCLUDED | WASTE HEAT RECOVERY | | VOL. PPM | LB/HR |
| 4A | 5 | 2 | Recycle Selectox | No | Yes | No | 2.45 | 6,540 | 101 |
| 6A | 10 | 2 | Claus | No | Yes | No | 3.53 | 6,330 | 194 |
| 6C | 10 | 2 | Recycle Selectox | No | Yes | No | 2.69 | 6,540 | 201 |
| 8A | 10 | 3 | Claus | No | Yes | No | 3.79 | 4,830 | 148 |
| 8C | 10 | 3 | Recycle Selectox | No | Yes | No | 3.16 | 4,240 | 130 |
| 10A | 100 | 3 | Claus | No | Yes | Yes | 11.21 | 4,830 | 1,480 |
| 10C | 100 | 3 | Recycle Selectox | No | Yes | Yes | 9.91 | 4,240 | 1,300 |

Table 5-4 (Table E-11)

## Onshore Sour Natural Gas Sulfur Recovery Study

## Summary of Investment Costs by Units (MM$)

| ACID GAS $H_2S/CO_2$ RATIO | 50/50 | | | | | | | | | | | | | | 80/20 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| CASE NO. | 1 | 3 | 5 | 7 | 9 | 11 | 13 | 15 | 17 | 20 | 23 | 26 | 29 | 32 | 19 |
| SULFUR INPUT (LT/D) | 5 | 5 | 10 | 10 | 100 | 100 | 100 | 100 | 100 | 1,000 | 1,000 | 1,000 | 1,000 | 1,000 | 1,000 |
| **SULFUR RECOVERY PROCESSES:** | | | | | | | | | | | | | | | |
| Recycle Selectox, 2-Stage | | 1.93 | | | | | | | | | | | | | |
| Claus, 2-Stage | | | 2.45 | | | | | | 3.73 | | | | | 14.50 | |
| Claus, 3-Stage | | | | 2.82 | 4.53 | 4.85 | 4.53 | 4.53 | | 17.73 | 19.49 | 17.73 | 17.73 | | 15.28 |
| **TAIL GAS PROCESSES:** | | | | | | | | | | | | | | | |
| BSR | | | | | | 1.66 | 1.58 | | | | 5.55 | 5.22 | | | |
| MDEA | | | | | | 4.07 | | | | | 17.57 | | | | |
| Stretford | | | | | | | 3.01 | | | | | 7.04 | | | |
| BSR (for Selectox) | | | | | | | | | 1.42 | | | | | 4.96 | |
| Selectox | | | | | | | | | 0.84 | | | | | 3.27 | |
| Sulfreen | | | | | | | | 2.21 | | | | | 5.26 | | |
| **THERMAL OXIDIZERS, WASTE HEAT BOILERS AND STACKS:** | | | | | | | | | | | | | | | |
| Thermal Oxidizer | 0.23 | 0.23 | 0.23 | 0.23 | 0.45 | 0.46 | | 0.45 | 0.45 | 1.89 | 1.94 | | 1.89 | 1.83 | 1.50 |
| Waste Heat Boiler(s) | | | | | 0.59 | 0.59 | | 0.61 | 0.57 | 2.35 | 2.35 | | 2.38 | 2.22 | 1.94 |
| Stack | 1.05 | 0.27 | 0.27 | 0.27 | 0.90 | 0.27 | | 0.46 | 0.46 | 4.13 | 0.27 | | 2.15 | 2.15 | 3.58 |
| **TOTAL INVESTMENT** | 1.28 | 2.43 | 2.95 | 3.32 | 6.47 | 11.90 | 9.12 | 8.26 | 7.47 | 26.10 | 47.17 | 29.99 | 29.41 | 28.93 | 22.30 |

## Table 5-5 (Table E-12)

## Onshore Sour Natural Gas Sulfur Recovery Study

## Summary of Investment Costs by Units (MM$)

| ACID GAS $H_2S/CO_2$ RATIO | 20/80 | | | | | | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| CASE NO. | 2 | 4 | 6 | 6B | 8 | 8B | 10 | 10B | 12 | 14 | 16 | 18 | 21 | 24 | 27 | 30 | 33 |
| SULFUR INPUT (LT/D) | 5 | 5 | 10 | 10 | 10 | 10 | 100 | 100 | 100 | 100 | 100 | 100 | 555 | 555 | 555 | 555 | 555 |
| **SULFUR RECOVERY PROCESSES:** | | | | | | | | | | | | | | | | | |
| Recycle Selectox, 2-Stage | | 1.93 | | 2.08 | | | | | | | | | | | | | |
| Claus, 2-Stage | | | 2.81 | | | | | | | | | 5.07 | | | | | 14.60 |
| Recycle Selectox, 3-Stage | | | | | | 2.61 | | 5.78 | | | | | | | | | |
| Claus, 3-Stage | | | | | 3.24 | | 6.23 | | 7.31 | 6.23 | 6.23 | | 17.93 | 21.39 | 17.93 | 17.93 | |
| **TAIL GAS PROCESSES:** | | | | | | | | | | | | | | | | | |
| BSR | | | | | | | | | 2.25 | 1.99 | | | | 6.01 | 5.22 | | |
| MDEA | | | | | | | | | 8.86 | | | | | 33.51 | | | |
| Stretford | | | | | | | | | | 3.31 | | | | | 6.27 | | |
| BSR (for Selectox) | | | | | | | | | | | | 1.80 | | | | | 4.93 |
| Selectox | | | | | | | | | | | | 1.33 | | | | | 3.79 |
| Sulfreen | | | | | | | | | | | 2.42 | | | | | 5.23 | |
| **THERMAL OXIDIZERS, WASTE HEAT BOILERS AND STACKS:** | | | | | | | | | | | | | | | | | |
| Thermal Oxidizer | 0.23 | 0.23 | 0.23 | 0.23 | 0.23 | 0.23 | 0.70 | 0.70 | 0.77 | | 0.70 | 0.72 | 2.18 | 2.38 | | 2.16 | 2.22 |
| Waste Heat Boiler(s) | | | | | | | 0.90 | 0.93 | 0.98 | | 0.92 | 0.92 | 2.54 | 2.79 | | 2.56 | 2.60 |
| Stack | 1.05 | 0.27 | 0.36 | 0.36 | 0.27 | 0.27 | 1.22 | 1.05 | 0.27 | | 0.61 | 0.61 | 3.58 | 0.27 | | 1.75 | 1.75 |
| **TOTAL INVESTMENT** | 1.28 | 2.43 | 3.40 | 2.67 | 3.74 | 3.11 | 9.05 | 8.46 | 20.44 | 11.53 | 10.88 | 10.45 | 26.23 | 66.35 | 29.42 | 29.63 | 29.89 |

Table 5-6 (Table E-13)

## Onshore Sour Natural Gas Sulfur Recovery Study

## Summary of Investment Costs by Units (MM$)

| ACID GAS $H_2S/CO_2$ RATIO | 12.5/87.5 | | | | | | |
|---|---|---|---|---|---|---|---|
| CASE NO. | 4A | 6A | 6C | 8A | 8C | 10A | 10C |
| SULFUR INPUT (LT/D) | 5 | 10 | 10 | 10 | 10 | 100 | 100 |
| **SULFUR RECOVERY PROCESSES:** | | | | | | | |
| Recycle Selectox, 2-Stage | 1.95 | | 2.10 | | | | |
| Claus, 2-Stage | | 2.94 | | | | | |
| Recycle Selectox, 3-Stage | | | | | 2.66 | | 6.60 |
| Claus, 3-Stage | | | | 3.29 | | 7.77 | |
| **THERMAL OXIDIZERS, WASTE HEAT BOILERS, AND STACKS:** | | | | | | | |
| Thermal Oxidizer | 0.23 | 0.23 | 0.23 | 0.23 | 0.23 | 0.91 | 0.91 |
| Waste Heat Boiler(s) | | | | | | 1.15 | 1.18 |
| Stack | 0.27 | 0.36 | 0.36 | 0.27 | 0.27 | 1.38 | 1.22 |
| TOTAL INVESTMENT | 2.45 | 3.53 | 2.69 | 3.79 | 3.16 | 11.21 | 9.91 |

## COST OF INITIAL CHARGE OF CATALYSTS AND CHEMICALS

The costs for providing the required amount of catalyst and chemicals to fill the plants for initial startup are provided in Tables 5-7, 5-8, and 5-9.

Table 5-7 (Table E-14)

## Onshore Sour Natural Gas Sulfur Recovery Study

## Cost of Initial Charge of Catalysts and Chemicals

## Acid Gas $H_2S/CO_2$ Ratios of 80/20 and 50/50

| CASE NO. | SULFUR INPUT (LT/D) | SULFUR RECOVERY UNIT NO. STAGES | SULFUR RECOVERY UNIT TYPE | TAIL GAS UNIT | THERMAL OXIDIZER INCLUDED | THERMAL OXIDIZER WASTE HEAT RECOVERY | CAT. & CHEM. COST (MM$) |
|---|---|---|---|---|---|---|---|
| **Acid Gas $H_2S/CO_2$ Ratio - 80/20** | | | | | | | |
| 19 | 1,000 | 3 | | No | Yes | Yes | 0.19 |
| **Acid Gas $H_2S/CO_2$ Ratio - 50/50** | | | | | | | |
| 1 | 5 | - | No | No | Yes | No | -0- |
| 3 | 5 | 2 | Recycle Selectox | No | Yes | No | 0.02 |
| 5 | 10 | 2 | Claus | No | Yes | No | 0.004 |
| 7 | 10 | 3 | Claus | No | Yes | No | 0.004 |
| 9 | 100 | 3 | Claus | No | Yes | Yes | 0.02 |
| 11 | 100 | 3 | Claus | BSR/MDEA | Yes | Yes | 0.13 |
| 13 | 100 | 3 | Claus | BSRP | No* | No | 0.10 |
| 15 | 100 | 3 | Claus | Sulfreen | Yes | Yes | 0.05 |
| 17 | 100 | 2 | Claus | BSR/Selectox | Yes | Yes | 0.16 |
| 20 | 1,000 | 3 | Claus | No | Yes | Yes | 0.25 |
| 23 | 1,000 | 3 | Claus | BSR/MDEA | Yes | Yes | 1.33 |
| 26 | 1,000 | 3 | Claus | BSRP | No* | No | 0.99 |
| 29 | 1,000 | 3 | Claus | Sulfreen | Yes | Yes | 0.50 |
| 32 | 1,000 | 2 | Claus | BSR/Selectox | Yes | Yes | 1.33 |

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*Combustor only

Table 5-8 (Table E-15)

## Onshore Sour Natural Gas Sulfur Recovery Study

## Cost of Initial Charge of Catalysts and Chemicals

## Acid Gas $H_2S/CO_2$ Ratio of 20/80

| CASE NO. | SULFUR INPUT (LT/D) | SULFUR RECOVERY UNIT |  | TAIL GAS UNIT | THERMAL OXIDIZER |  | CAT. & CHEM. COST (MM$) |
|---|---|---|---|---|---|---|---|
|  |  | NO. STAGES | TYPE |  | INCLUDED | WASTE HEAT RECOVERY |  |
| 2 | 5 | - | No | No | Yes | No | -0- |
| 4 | 5 | 2 | Recycle Selectox | No | Yes | No | 0.01 |
| 6 | 10 | 2 | Claus | No | Yes | No | 0.004 |
| 6B | 10 | 2 | Recycle Selectox | No | Yes | No | 0.03 |
| 8 | 10 | 3 | Claus | No | Yes | No | 0.004 |
| 8B | 10 | 3 | Recycle Selectox | No | Yes | No | 0.03 |
| 10 | 100 | 3 | Claus | No | Yes | Yes | 0.05 |
| 10B | 100 | 3 | Recycle Selectox | No | Yes | Yes | 0.30 |
| 12 | 100 | 3 | Claus | BSR/MDEA | Yes | Yes | 0.36 |
| 14 | 100 | 3 | Claus | BSRP | No* | No | 0.17 |
| 16 | 100 | 3 | Claus | Sulfreen | Yes | Yes | 0.09 |
| 18 | 100 | 2 | Claus | BSR/Selectox | Yes | Yes | 0.27 |
| 21 | 555 | 3 | Claus | No | Yes | Yes | 0.25 |
| 24 | 555 | 3 | Claus | BSR/MDEA | Yes | Yes | 2.01 |
| 27 | 555 | 3 | Claus | BSRP | No* | No | 0.94 |
| 30 | 555 | 3 | Claus | Sulfreen | Yes | Yes | 0.50 |
| 33 | 555 | 2 | Claus | BSR/Selectox | Yes | Yes | 0.87 |

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*Combustor only

Table 5-9 (Table E-16)

Onshore Sour Natural Gas Sulfur Recovery Study

Cost of Initial Charge of Catalysts and Chemicals

Acid Gas $H_2S/CO_2$ Ratio of 12.5/87.5

| CASE NO. | SULFUR INPUT (LT/D) | SULFUR RECOVERY UNIT | | TAIL GAS UNIT | THERMAL OXIDIZER | | CAT. & CHEM. COST (MM$) |
|---|---|---|---|---|---|---|---|
| | | NO. STAGES | TYPE | | INCLUDED | WASTE HEAT RECOVERY | |
| 4A | 5 | 2 | Recycle Selectox | No | Yes | No | 0.01 |
| 6A | 10 | 2 | Claus | No | Yes | No | 0.004 |
| 6C | 10 | 2 | Recycle Selectox | No | Yes | No | 0.03 |
| 8A | 10 | 3 | Claus | No | Yes | No | 0.01 |
| 8C | 10 | 3 | Recycle Selectox | No | Yes | No | 0.03 |
| 10A | 100 | 3 | Claus | No | Yes | Yes | 0.06 |
| 10C | 100 | 3 | Recycle Selectox | No | Yes | Yes | 0.31 |

## SECTION 6 (SECTION E-6)

## DIRECT OPERATING COST DATA

### UTILITIES, CATALYSTS, AND CHEMICALS

The utilities and the catalyst and chemical costs are provided in Tables 6-1 through 6-8 for all cases except No. 1 and No. 2, which do not require any.

The utilities for each case are in two groups, those to be supplied to battery limits, which are costs, and those exported to battery limits, which are credits.

The Claus units produce 250-, 50-, and 15-psig steam. The 15-psig steam is sometimes condensed with the energy being wasted but the condensate being recovered. However, it is exported to battery limits in all cases as it can be useful in preheating or deaerating boiler feedwater.

The waste heat boilers recovering heat by cooling hydrogenation reactor effluent gas produce 50-psig steam for the 100-LT/D cases and both 450- and 50-psig steam for the 555- and 1,000-LT/D cases.

The thermal oxidizer waste heat boilers produce 250-psig steam for the 100-LT/D cases and both 600- and 250-psig steam for the 555- and 1,000-LT/D cases.

Chemical costs for the Stretford process, in industrial experience, have varied widely. Plant design has a major effect on such costs, especially the design of the sulfur slurry separation section of the plant. Both investment and chemical costs herein apply to advanced design incorporating very complete separations, with relatively high investment costs but low chemical costs. It will be noted that a relatively large error in estimating chemical costs has but a minor effect on overall costs.

Table 6-1 (Table E-17)

**Onshore Sour Natural Gas Sulfur Recovery Study**

**Utilities and Catalyst Costs**

**Recycle Selectox 2-Stage Process**

| CASE NO. | 3 | 4 | 4A | 6B | 6C |
|---|---|---|---|---|---|
| Sulfur Input, LT/D | 5 | 5 | 5 | 10 | 10 |
| $H_2S/CO_2$ Ratio | 50/50 | 20/80 | 12.5/87.5 | 20/80 | 12.5/87.5 |
| REQUIRED FROM BATTERY LIMITS | | | | | |
| 600-psig Steam, lb/hr | 373 | 616 | 857 | 1,230 | 1,713 |
| Treated B.F. Water, lb/hr | 1,566 | 1,652 | 1,740 | 3,304 | 3,379 |
| Electric Power, kW | 77 | 70 | 65 | 110 | 100 |
| Fuel Gas, $10^6$ Btu/hr | 0.59 | 1.22 | 1.88 | 2.44 | 3.76 |
| EXPORTED TO BATTERY LIMITS | | | | | |
| 50-psig Steam, lb/hr | 635 | 660 | 675 | 1,920 | 1,950 |
| 15-psig Steam, lb/hr | 76 | 144 | 214 | 288 | 429 |
| Steam Condensate, lb/hr | 1,173 | 1,416 | 2,275 | 2,230 | 3,950 |
| Catalyst Cost, $/day | 4.73 | 4.70 | 4.97 | 9.40 | 9.93 |

Table 6-2 (Table E-18)

## Onshore Sour Natural Gas Sulfur Recovery Study

## Utilities and Catalyst Costs

## Claus Process - No Tail Gas or Waste Heat Recovery Units

| CASE NO. | 5 | 6 | 6A | 7 | 8 | 8A |
|---|---|---|---|---|---|---|
| Claus - No. Stages | 2 | 2 | 2 | 3 | 3 | 3 |
| Sulfur Input, LT/D | 10 | 10 | 10 | 10 | 10 | 10 |
| $H_2S/CO_2$ Ratio | 50/50 | 20/80 | 12.5/87.5 | 50/50 | 20/80 | 12.5/87.5 |
| REQUIRED FROM BATTERY LIMITS | | | | | | |
| 600-psig Steam, lb/hr | 239 | 848 | 1,302 | 343 | 1,057 | 1,615 |
| Treated B.F. Water, lb/hr | 2,560 | 2,894 | 3,095 | 2,636 | 3,044 | 3,308 |
| Electric Power, kW | 51 | 56 | 59 | 53 | 58 | 61 |
| Fuel Gas, $10^6$ x Btu/hr | 0.95 | 2.34 | 3.72 | 1.01 | 2.43 | 3.78 |
| EXPORTED TO BATTERY LIMITS | | | | | | |
| 250-psig Steam, lb/hr | 1,409 | 1,277 | 1,239 | 1,409 | 1,277 | 1,239 |
| 50-psig Steam, lb/hr | 0 | 122 | 221 | 129 | 364 | 512 |
| 15-psig Steam, lb/hr | 217 | 411 | 545 | 162 | 314 | 461 |
| Steam Condensate, lb/hr | 1,096 | 1,848 | 2,302 | 1,202 | 2,057 | 2,615 |
| Catalyst Cost, $/day | 2.27 | 4.12 | 5.86 | 3.40 | 6.18 | 8.78 |

Table 6-3 (Table E-19)

## Onshore Sour Natural Gas Sulfur Recovery Study

## Utilities and Catalyst Costs

## Claus Process - With Waste Heat Recovery - No Tail Gas Unit

| CASE NO. | 9 | 10 | 10A | 19 | 20 | 21 |
|---|---|---|---|---|---|---|
| Claus - No. Stages | 3 | 3 | 3 | 3 | 3 | 3 |
| Sulfur Input, LT/D | 100 | 100 | 100 | 1,000 | 1,000 | 555 |
| $H_2S/CO_2$ Ratio | 50/50 | 20/80 | 12.5/87.5 | 80/20 | 50/50 | 20/80 |
| **REQUIRED FROM BATTERY LIMITS** | | | | | | |
| Treated B.F. Water, lb/hr | 36,650 | 51,770 | 65,230 | 332,000 | 365,000 | 285,300 |
| Electric Power, kW | 233 | 273 | 308 | 1,676 | 1,876 | 1,275 |
| Fuel Gas, $10^6$ x Btu/hr | 10.11 | 24.30 | 37.84 | 83.13 | 124.6 | 159.3 |
| **EXPORTED TO BATTERY LIMITS** | | | | | | |
| 600-psig Steam, lb/hr | 5,490 | 7,930 | 11,710 | 65,305 | 89,200 | 44,036 |
| 250-psig Steam, lb/hr | 15,160 | 14,984 | 15,736 | 169,206 | 151,601 | 83,165 |
| 50-psig Steam, lb/hr | 7,880 | 11,640 | 12,120 | 79,751 | 104,211 | 82,280 |
| 15-psig Steam, lb/hr | 1,620 | 3,140 | 4,610 | 4,043 | 5,400 | 5,800 |
| Steam Condensate, lb/hr | 5,430 | 12,570 | 19,150 | 4,000 | 4,000 | 61,664 |
| Catalyst Cost, $/day | 34.00 | 61.80 | 87.83 | 266.98 | 339.98 | 343.08 |

Table 6-4 (Table E-20)

## Onshore Sour Natural Gas Sulfur Recovery Study

## Utilities and Catalyst and Chemical Costs

## BSR/MDEA Tail Gas Cases

| CASE NO. | 11 | 12 | 23 | 24 |
|---|---|---|---|---|
| Claus - No. Stages | 3 | 3 | 3 | 3 |
| Sulfur Input, LT/D | 100 | 100 | 1,000 | 555 |
| $H_2S/CO_2$ Ratio | 50/50 | 20/80 | 50/50 | 20/80 |
| REQUIRED FROM BATTERY LIMITS | | | | |
| Cooling Water (25°F Rise), gpm | 1,638 | 6,021 | 16,380 | 33,420 |
| 50-psig Steam Required, lb/hr | 13,277 | 77,523 | 133,904 | 436,073 |
| Treated B.F. Water, lb/hr | 44,430 | 70,950 | 442,500 | 391,500 |
| Electric Power, kW | 565 | 1,020 | 4,513 | 5,318 |
| Fuel Gas, $10^6$ x Btu/hr | 20.5 | 48.0 | 205 | 266.5 |
| EXPORTED TO BATTERY LIMITS | | | | |
| 600-psig Steam, lb/hr | 5,213 | 7,747 | 89,600 | 72,542 |
| 450-psig Steam, lb/hr | - | - | 28,590 | 8,824 |
| 250-psig Steam, lb/hr | 15,406 | 15,865 | 154,056 | 88,051 |
| 15-psig Steam, lb/hr | 1,950 | 4,202 | 6,498 | 7,770 |
| Steam Condensate, lb/hr | 33,547 | 117,663 | 281,800 | 644,976 |
| Catalyst and Chemical Costs, $/day | 292.40 | 645.00 | 2,923.90 | 3,578.60 |

Table 6-5 (Table E-21)

## Onshore Sour Natural Gas Sulfur Recovery Study

## Utilities and Catalyst Costs

## Recycle Selectox 3-Stage Process

| CASE NO. | 8B | 8C | 10B | 10C |
|---|---|---|---|---|
| Sulfur Input, LT/D | 10 | 10 | 100 | 100 |
| $H_2S/CO_2$ Ratio | 20/80 | 12.5/87.5 | 20/80 | 12.5/87.5 |
| Waste Heat Recovery | No | No | Yes | Yes |
| **REQUIRED FROM BATTERY LIMITS** | | | | |
| 600-psig Steam, lb/hr | 1,538 | 2,141 | | |
| Treated B.F. Water, lb/hr | 3,551 | 3,782 | 57,038 | 70,148 |
| Electric Power, kW | 114 | 104 | 827 | 727 |
| Fuel Gas, $10^6$ Btu/hr | 2.54 | 3.82 | 25.37 | 38.18 |
| **EXPORTED TO BATTERY LIMITS** | | | | |
| 600-psig Steam, lb/hr | | | 3,275 | 6,610 |
| 250-psig Steam, lb/hr | | | 2,232 | 3,365 |
| 50-psig Steam, lb/hr | 2,160 | 2,240 | 36,100 | 30,930 |
| 15-psig Steam, lb/hr | 288 | 429 | 2,880 | 4,290 |
| Steam Condensate, lb/hr | 2,538 | 3,243 | 16,880 | 32,910 |
| Catalyst Cost, $/day | 11.37 | 12.76 | 114.67 | 128.04 |

Table 6-6 (Table E-22)

## Onshore Sour Natural Gas Sulfur Recovery Study

## Utilities and Catalyst and Chemical Costs

## Beavon Sulfur Removal Process (BSRP)

| CASE NO. | 13 | 14 | 26 | 27 |
|---|---|---|---|---|
| Sulfur Input, LT/D | 100 | 100 | 1,000 | 555 |
| $H_2S/CO_2$ Ratio | 50/50 | 20/80 | 50/50 | 20/80 |
| REQUIRED FROM BATTERY LIMITS | | | | |
| Treated B.F. Water, lb/hr | 31,240 | 39,920 | 310,400 | 219,500 |
| Cooling Water (25°F Rise), gpm | 1,320 | 2,136 | 13,200 | 11,850 |
| Electric Power, kW | 700 | 870 | 5,850 | 4,530 |
| Fuel Gas, $10^6$ x Btu/hr | 7.34 | 17.65 | 73.36 | 97.97 |
| EXPORTED TO BATTERY LIMITS | | | | |
| 450-psig Steam, lb/hr | - | - | 24,030 | 27,420 |
| 250-psig Steam, lb/hr | 14,580 | 10,230 | 145,800 | 56,780 |
| 50-psig Steam, lb/hr | 11,770 | 20,020 | 118,460 | 101,490 |
| 15-psig Steam, lb/hr | 1,380 | 2,400 | 3,020 | 1,670 |
| Steam Condensate, lb/hr | 2,360 | 5,364 | 7,650 | 21,645 |
| Catalyst and Chemical Costs, $/day | 181.40 | 295.30 | 1,814.00 | 1,639.00 |

Table 6-7 (Table E-23)

## Onshore Sour Natural Gas Sulfur Recovery Study

## Utilities and Catalyst Costs

## BSR/Selectox Process

| CASF NO. | 17 | 18 | 32 | 33 |
|---|---|---|---|---|
| Claus - No. Stages | 2 | 2 | 2 | 2 |
| Sulfur Input, LT/D | 100 | 100 | 1,000 | 555 |
| $H_2S/CO_2$ Ratio | 50/50 | 20/80 | 50/50 | 20/80 |
| **REQUIRED FROM BATTERY LIMITS** | | | | |
| Cooling Water (25°F Rise), gpm | 186 | 388 | 1,860 | 2,153 |
| Treated B.F. Water, lb/hr | 39,383 | 60,028 | 393,201 | 333,282 |
| Electric Power, kW | 343 | 419 | 2,575 | 1,913 |
| Fuel Gas, $10^6$ x Btu/hr | 15.3 | 37.2 | 153 | 206 |
| **EXPORTED TO BATTERY LIMITS** | | | | |
| 600-psig Steam, lb/hr | 8,200 | 19,178 | 82,003 | 106,455 |
| 450-psig Steam, lb/hr | - | - | 23,890 | 27,300 |
| 250-psig Steam, lb/hr | 15,066 | 15,052 | 150,656 | 83,539 |
| 50-psig Steam, lb/hr | 12,800 | 19,940 | 119,800 | 100,481 |
| 15-psig Steam, lb/hr | 2,170 | 4,110 | 5,400 | 5,800 |
| Steam Condensate, lb/hr | 7,227 | 17,007 | 32,370 | 86,288 |
| Catalyst Cost, $/day | 101.00 | 195.00 | 1,010.00 | 1,082.00 |

Table 6-8 (Table E-24)

### Onshore Sour Natural Gas Sulfur Recovery Study

### Utilities and Catalyst Costs

### Sulfreen Process

| CASE NO. | 15 | 16 | 29 | 30 |
|---|---|---|---|---|
| Claus - No. Stages | 3 | 3 | 3 | 3 |
| Sulfur Input, LT/D | 100 | 100 | 1,000 | 555 |
| $H_2S/CO_2$ Ratio | 50/50 | 20/80 | 50/50 | 20/80 |
| REQUIRED FROM BATTERY LIMITS | | | | |
| Treated B.F. Water, lb/hr | 37,348 | 55,737 | 372,061 | 307,234 |
| Electric Power, kW | 365 | 445 | 2,863 | 2,130 |
| Fuel Gas, $10^6$ x Btu/hr | 11.9 | 26.9 | 119.4 | 149.5 |
| EXPORTED TO BATTERY LIMITS | | | | |
| 600-psig Steam, lb/hr | 6,106 | 11,382 | 95,367 | 63,172 |
| 250-psig Steam, lb/hr | 15,224 | 15,382 | 152,246 | 85,369 |
| 50-psig Steam, lb/hr | 7,669 | 11,273 | 102,100 | 80,241 |
| 15-psig Steam, lb/hr | 1,620 | 3,140 | 5,400 | 5,800 |
| Steam Condensate, lb/hr | 5,641 | 12,937 | 6,111 | 63,703 |
| Catalyst Cost, $/day | 68.00 | 124.00 | 680.00 | 686.00 |

## SULFUR RECOVERED

The average sulfur recovered in LT/D for each of the cases is given in Tables 6-9, 6-10, and 6-11. These figures are based on the average of start-of-run and end-of-run recoveries, and are provided for applying credits for sulfur sales in the calculation of operating costs.

Table 6-9 (Table E-25)

Onshore Sour Natural Gas Sulfur Recovery Study

Sulfur Recovered (LT/D), Average During Run

Acid Gas $H_2S/CO_2$ Ratios of 80/20 and 50/50

| CASE NO. | SULFUR INPUT (LT/D) | SULFUR RECOVERY UNIT NO. STAGES | SULFUR RECOVERY UNIT TYPE | TAIL GAS UNIT | THERMAL OXIDIZER INCLUDED | THERMAL OXIDIZER WASTE HEAT RECOVERY | SULFUR RECOVERED (LT/D) |
|---|---|---|---|---|---|---|---|
| **Acid Gas $H_2S/CO_2$ Ratio - 80/20** | | | | | | | |
| 19 | 1,000 | 3 | | No | Yes | Yes | 966.40 |
| **Acid Gas $H_2S/CO_2$ Ratio - 50/50** | | | | | | | |
| 1 | 5 | - | No | No | Yes | No | -0- |
| 3 | 5 | 2 | Recycle Selectox | No | Yes | No | 4.69 |
| 5 | 10 | 2 | Claus | No | Yes | No | 9.38 |
| 7 | 10 | 3 | Claus | No | Yes | No | 9.61 |
| 9 | 100 | 3 | Claus | No | Yes | Yes | 96.05 |
| 11 | 100 | 3 | Claus | BSR/MDEA | Yes | Yes | 99.95 |
| 13 | 100 | 3 | Claus | BSRP | No* | No | 99.99 |
| 15 | 100 | 3 | Claus | Sulfreen | Yes | Yes | 98.42 |
| 17 | 100 | 2 | Claus | BSR/Selectox | Yes | Yes | 98.49 |
| 20 | 1,000 | 3 | Claus | No | Yes | Yes | 960.50 |
| 23 | 1,000 | 3 | Claus | BSR/MDEA | Yes | Yes | 999.50 |
| 26 | 1,000 | 3 | Claus | BSRP | No* | No | 999.90 |
| 29 | 1,000 | 3 | Claus | Sulfreen | Yes | Yes | 984.20 |
| 32 | 1,000 | 2 | Claus | BSR/Selectox | Yes | Yes | 984.90 |

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*Combustor only

Table 6-10 (Table E-26)

**Onshore Sour Natural Gas Sulfur Recovery Study**

**Sulfur Recovered (LT/D), Average During Run**

**Acid Gas $H_2S/CO_2$ Ratio of 20/80**

| CASE NO. | SULFUR INPUT (LT/D) | SULFUR RECOVERY UNIT NO. STAGES | SULFUR RECOVERY UNIT TYPE | TAIL GAS UNIT | THERMAL OXIDIZER INCLUDED | THERMAL OXIDIZER WASTE HEAT RECOVERY | SULFUR RECOVERED (LT/D) |
|---|---|---|---|---|---|---|---|
| 2 | 5 | - | No | No | Yes | No | -0- |
| 4 | 5 | 2 | Recycle Selectox | No | Yes | No | 4.60 |
| 6 | 10 | 2 | Claus | No | Yes | No | 9.18 |
| 6B | 10 | 2 | Recycle Selectox | No | Yes | No | 9.20 |
| 8 | 10 | 3 | Claus | No | Yes | No | 9.45 |
| 8B | 10 | 3 | Recycle Selectox | No | Yes | No | 9.54 |
| 10 | 100 | 3 | Claus | No | Yes | Yes | 94.50 |
| 10B | 100 | 3 | Recycle Selectox | No | Yes | Yes | 95.39 |
| 12 | 100 | 3 | Claus | BSR/MDEA | Yes | Yes | 99.90 |
| 14 | 100 | 3 | Claus | BSRP | No* | No | 99.99 |
| 16 | 100 | 3 | Claus | Sulfreen | Yes | Yes | 97.80 |
| 18 | 100 | 2 | Claus | BSR/Selectox | Yes | Yes | 97.84 |
| 21 | 555 | 3 | Claus | No | Yes | Yes | 524.50 |
| 24 | 555 | 3 | Claus | BSR/MDEA | Yes | Yes | 554.40 |
| 27 | 555 | 3 | Claus | BSRP | No* | No | 554.90 |
| 30 | 555 | 3 | Claus | Sulfreen | Yes | Yes | 542.80 |
| 33 | 555 | 2 | Claus | BSR/Selectox | Yes | Yes | 543.00 |

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*Combustor only

Table 6-11 (Table E-27)

Onshore Sour Natural Gas Sulfur Recovery Study

Sulfur Recovered (LT/D), Average During Run

Acid Gas $H_2S/CO_2$ Ratio of 12.5/87.5

| CASE NO. | SULFUR INPUT (LT/D) | SULFUR RECOVERY UNIT | | TAIL GAS UNIT | THERMAL OXIDIZER | | SULFUR RECOVERED (LT/D) |
|---|---|---|---|---|---|---|---|
| | | NO. STAGES | TYPE | | INCLUDED | WASTE HEAT RECOVERY | |
| 4A | 5 | 2 | Recycle Selectox | No | Yes | No | 4.55 |
| 6A | 10 | 2 | Claus | No | Yes | No | 9.13 |
| 6C | 10 | 2 | Recycle Selectox | No | Yes | No | 9.10 |
| 8A | 10 | 3 | Claus | No | Yes | No | 9.34 |
| 8C | 10 | 3 | Recycle Selectox | No | Yes | No | 9.42 |
| 10A | 100 | 3 | Claus | No | Yes | Yes | 93.39 |
| 10C | 100 | 3 | Recycle Selectox | No | Yes | Yes | 94.19 |

## OTHER DIRECT OPERATING COST DATA

(1) The required number of operators per shift are:

| Case No. | No. Per Shift |
|---|---|
| 1 and 2 | Negligible |
| 3, 4, 4A, 6B, 6C, 8B, 8C, 10B, 10C | 0.75 |
| 5, 6, 6A, 7, 8, 8A, 9, 10, 10A, 19, 20, 21 | 1.25 |
| All others | 2.25 |

Add 0.25 to all cases except 1 and 2 for shift supervision.

(2) Values for utilities

Steam:

| Pressure, PSIG | 600 | 450 | 250 | 50 | 15 |
|---|---|---|---|---|---|
| $/1,000 lbs | 7.25 | 7.00 | 6.75 | 5.75 | 4.50 |

| | |
|---|---|
| Treated boiler feed water | $1.50/1,000 lbs |
| Recovered condensate | $1.25/1,000 lbs |
| Cooling water | $0.05/1,000 circulated gas |
| Fuel | $5.00/MMBtu's |
| Electricity | $0.05/KW hr |
| Credit for recovered sulfur | $100/LT |

(3) For annual maintenance costs assume:

| | |
|---|---|
| Labor | 1½% of investment |
| Materials | 2% of investment |
| Operating supplies | 10% of operating labor |
| Overhead | 25% of operating and maintenance labor |
| Depreciation | 20 years life |
| Insurance | 0.6% of investment |

(4) Operating schedule: 24 h/d, 350 d/y.

## CAPITAL COSTS

The fixed-capital costs that represent the initial investment for control equipment and installation (including catalyst charge) for 39 different cases, which represent the size and acid gas ratio ranges in the industry, are presented in Table 6-12 (Table E-28). The cases consider different model plant sizes and combinations of sulfur recovery/tail gas processes and acid gas $H_2S/CO_2$ ratios.

## ANNUALIZED COSTS

The annualized costs for 39 different cases, which represent the size and acid gas ratio ranges in the industry, are presented in Table 6-13 (Table E-29). The cases are the same as those for which capital costs were developed in Table E-28.

## Table E-28. FIXED-CAPITAL COSTS FOR 39 CASES OF SULFUR INTAKE/ACID GAS $H_2S/CO_2$ RATIO/SULFUR RECOVERY TECHNOLOGY COMBINATIONS[a]

| Case No.[b] | Size (sulfur intake) Mg/d | Size (sulfur intake) LT/D | Acid gas feed $H_2S/CO_2$ ratio by volume | Sulfur recovery technology | Cost $ | Case No. | Size (sulfur intake) Mg/d | Size (sulfur intake) LT/D | Acid gas feed $H_2S/CO_2$ ratio by volume | Sulfur recovery technology | Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 1(1) | 5.1 | 5 | 50/50 | No recovery | 1,280,000 | 21(11) | 101.6 | 100 | 50/50 | Claus-3 stage with SCOT/BSRMDEA | 12,030,000 |
| 2(2) | 5.1 | 5 | 20/80 | No recovery | 1,280,000 | 22(12) | 101.6 | 100 | 20/80 | Claus-3 stage with SCOT/BSRMDEA | 20,800,000 |
| 3(3) | 5.1 | 5 | 50/50 | Recycle Selectox-2 stage | 2,450,000 | 23(13) | 101.6 | 100 | 50/50 | Claus-3 stage with BSRP | 9,220,000 |
| 4(4) | 5.1 | 5 | 20/80 | Recycle Selectox-2 stage | 2,440,000 | 24(14) | 101.6 | 100 | 20/80 | Claus-3 stage with BSRP | 11,700,000 |
| 5(4A) | 5.1 | 5 | 12.5/87.5 | Recycle Selectox-2 stage | 2,460,000 | 25(15) | 101.6 | 100 | 50/50 | Claus-3 stage with Sulfreen | 8,310,000 |
| 6(5) | 10.2 | 10 | 50/50 | Claus-2 stage | 2,950,000 | 26(16) | 101.6 | 100 | 20/80 | Claus-3 stage with Sulfreen | 10,970,000 |
| 7(6) | 10.2 | 10 | 20/80 | Claus-2 stage | 3,400,000 | 27(17) | 101.6 | 100 | 50/50 | Claus-2 stage with BSR/Selectox I | 7,630,000 |
| 8(6A) | 10.2 | 10 | 12.5/87.5 | Claus-2 stage | 3,530,000 | 28(18) | 101.6 | 100 | 20/80 | Claus-2 stage with BSR/Selectox I | 10,720,000 |
| 9(6B) | 10.2 | 10 | 20/80 | Recycle Selectox-2 stage | 2,700,000 | 29(19) | 1016.0 | 1000 | 80/20 | Claus-3 stage | 22,490,000 |
| 10(6C) | 10.2 | 10 | 12.5/87.5 | Recycle Selectox-2 stage | 2,720,000 | 30(20) | 1016.0 | 1000 | 50/50 | Claus-3 stage | 26,350,000 |
| 11(7) | 10.2 | 10 | 50/50 | Claus-3 stage | 3,320,000 | 31(21) | 563.9 | 555 | 20/80 | Claus-3 stage | 26,480,000 |
| 12(8) | 10.2 | 10 | 20/80 | Claus-3 stage | 3,740,000 | 32(23) | 1016.0 | 1000 | 50/50 | Claus-3 stage with SCOT/BSRMDEA | 48,500,000 |
| 13(8A) | 10.2 | 10 | 12.5/87.5 | Claus-3 stage | 3,800,000 | 33(24) | 563.9 | 555 | 20/80 | Claus-3 stage with SCOT/BSRMDEA | 68,360,000 |
| 14(8B) | 10.2 | 10 | 20/80 | Recycle Selectox-3 stage | 3,140,000 | 34(26) | 1016.0 | 1000 | 50/50 | Claus-3 stage with BSRP | 30,980,000 |
| 15(8C) | 10.2 | 10 | 12.5/87.5 | Recycle Selectox-3 stage | 3,190,000 | 35(27) | 563.9 | 555 | 20/80 | Claus-3 stage with BSRP | 30,360,000 |
| 16(9) | 101.6 | 100 | 50/50 | Claus-3 stage | 6,490,000 | 36(29) | 1016.0 | 1000 | 50/50 | Claus-3 stage with Sulfreen | 29,910,000 |
| 17(10) | 101.6 | 100 | 20/80 | Claus-3 stage | 9,100,000 | 37(30) | 563.9 | 555 | 20/80 | Claus-3 stage with Sulfreen | 30,130,000 |
| 18(10A) | 101.6 | 100 | 12.5/87.5 | Claus-3 stage | 11,270,000 | 38(32) | 1016.0 | 1000 | 50/50 | Claus-2 stage with BSR/Selectox I | 30,260,000 |
| 19(10B) | 101.6 | 100 | 20/80 | Recycle Selectox-3 stage | 8,760,000 | 39(33) | 563.9 | 555 | 20/80 | Claus-2 stage with BSR/Selectox I | 30,760,000 |
| 20(10C) | 101.6 | 100 | 12.5/87.5 | Recycle Selectox-3 stage | 10,220,000 | | | | | | |

[a] The costs were extracted from Tables E-8 through E-10 in Section E.5 of Appendix E.

[b] The case numbers in the parentheses are the numbers that were assigned when the study was performed.

## Table E-29. ANNUALIZED COSTS FOR 39 CASES OF SULFUR INTAKE/ACID GAS $H_2S/CO_2$ RATIO/SULFUR RECOVERY TECHNOLOGY COMBINATIONS[a]

| Case No.[b] | Size (sulfur intake) Mg/d | LT/D | Acid gas feed $H_2S/CO_2$ ratio by volume | Sulfur recovery technology | Cost $ | Case No. | Size (sulfur intake) Mg/d | LT/D | Acid gas feed $H_2S/CO_2$ ratio by volume | Sulfur recovery technology | Cost $ |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 1(1) | 5.1 | 5 | 50/50 | No recovery | 231,365 | 21(11) | 101.6 | 100 | 50/50 | Claus-3 stage with SCOT/BSRMDEA | 34,738 |
| 2(2) | 5.1 | 5 | 20/80 | No recovery | 231,365 | 22(12) | 101.6 | 100 | 20/80 | Claus-3 stage with SCOT/BSRMDEA | 5,590,552 |
| 3(3) | 5.1 | 5 | 50/50 | Recycle Selectox-2 stage | 537,876 | 23(13) | 101.6 | 100 | 50/50 | Claus-3 stage with BSRP | (1,707,122) |
| 4(4) | 5.1 | 5 | 20/80 | Recycle Selectox-2 stage | 573,067 | 24(14) | 101.6 | 100 | 20/80 | Claus-3 stage with BSRP | (793,079) |
| 5(4A) | 5.1 | 5 | 12.5/87.5 | Recycle Selectox-2 stage | 608,373 | 25(15) | 101.6 | 100 | 50/50 | Claus-3 stage with Sulfreen | (2,014,549) |
| 6(5) | 10.2 | 10 | 50/50 | Claus-2 stage | 512,149 | 26(16) | 101.6 | 100 | 20/80 | Claus-3 stage with Sulfreen | (1,217,247) |
| 7(6) | 10.2 | 10 | 20/80 | Claus-2 stage | 691,324 | 27(17) | 101.6 | 100 | 50/50 | Claus-2 stage with BSR/Selectox I | (2,364,802) |
| 8(6A) | 10.2 | 10 | 12.5/87.5 | Claus-2 stage | 795,924 | 28(18) | 101.6 | 100 | 20/80 | Claus-2 stage with BSR/Selectox I | (1,697,336) |
| 9(6B) | 10.2 | 10 | 20/80 | Recycle Selectox-2 stage | 514,668 | 29(19) | 1016.0 | 1000 | 80/20 | Claus-3 stage | (38,439,508) |
| 10(6C) | 10.2 | 10 | 12.5/87.5 | Recycle Selectox-2 stage | 580,363 | 30(20) | 1016.0 | 1000 | 50/50 | Claus-3 stage | (36,912,577) |
| 11(7) | 10.2 | 10 | 50/50 | Claus-3 stage | 576,982 | 31(21) | 563.9 | 555 | 20/80 | Claus-3 stage | (14,353,713) |
| 12(8) | 10.2 | 10 | 20/80 | Claus-3 stage | 753,471 | 32(23) | 1016.0 | 1000 | 50/50 | Claus-3 stage with SCOT/BSRMDEA | (20,345,209) |
| 13(8A) | 10.2 | 10 | 12.5/87.5 | Claus-3 stage | 849,862 | 33(24) | 563.9 | 555 | 20/80 | Claus-3 stage with SCOT/BSRMDEA | 18,787,646 |
| 14(8B) | 10.2 | 10 | 20/80 | Recycle Selectox-3 stage | 596,480 | 34(26) | 1016.0 | 1000 | 50/50 | Claus-3 stage with BSRP | (33,884,038) |
| 15(8C) | 10.2 | 10 | 12.5/87.5 | Recycle Selectox-3 stage | 684,596 | 35(27) | 563.9 | 555 | 20/80 | Claus-3 stage with BSRP | (13,625,880) |
| 16(9) | 101.6 | 100 | 50/50 | Claus-3 stage | (2,569,608) | 36(29) | 1016.0 | 1000 | 50/50 | Claus-3 stage with Sulfreen | (36,830,877) |
| 17(10) | 101.6 | 100 | 20/80 | Claus-3 stage | (1,666,958) | 37(30) | 563.9 | 555 | 20/80 | Claus-3 stage with Sulfreen | (15,009,103) |
| 18(10A) | 101.6 | 100 | 12.5/87.5 | Claus-3 stage | (463,637) | 38(32) | 1016.0 | 1000 | 50/50 | Claus-2 stage with BSR/Selectox I | (36,670,098) |
| 19(10B) | 101.6 | 100 | 20/80 | Recycle Selectox-3 stage | (1,694,062) | 39(33) | 563.9 | 555 | 20/80 | Claus-2 stage with BSR/Selectox I | (17,395,397) |
| 20(10C) | 101.6 | 100 | 12.5/87.5 | Recycle Selectox-3 stage | (948,202) | | | | | | |

[a] The costs determined from the data presented in Section E of Appendix E, and Chapter 8 References 2 and 3. Also refer to the Docket (A-80-20-A) entry (II-B-29) for the methodology and sample calculations.

[b] The case numbers in the parentheses are the numbers that were assigned when the study was performed.

# APPENDIX F

Unit natural gas production cost, UPC, (Table 9-4) equals:

$$UPC = \frac{\left[1\text{-}itc - t \sum_{y=1}^{n} \frac{Dep_y}{(1+d)^y}\right]\left[I_d + I_e\right] + (1-t)\ OM_o \sum_{y=1}^{L} \frac{(1+qrate)^y}{(1+r)^y}}{Q_o\ (1-roy)\ (1-t + t\ depl) \sum_{y=1}^{L} \frac{(1+qrate)^y}{(1+r)^y}}$$

It was assumed that:

itc = investment tax credit rate, .1

t = corporate income tax rate on marginal income, .47

n = depreciation period, 10

$Dep_y$ = ACRS depreciation rates, 10 year life

d = nominal discount factor, .10

$I_d$ = drilling cost, input $

$I_e$ = equipment cost, input $

$OC_o$ = annual operating and maintenance cost, input $

L = well life, 20

qrate = production (and operating cost) decline rate, -0.06

r = real discount factor, .019

$Q_o$ = annual production, input mcfy.

roy = royalty payment rate, .2

depl = percentage depletion allowance, .15

# APPENDIX G – THE AMERICAN PETROLEUM INSTITUTE'S GAS PLANT SURVEY DATA

The American Petroleum Institute (API) submitted a report on their gas plant survey containing data/information on 731 gas streams. The data included current (July 1982) capacity, current throughput, and gas stream $H_2S$ and $CO_2$ concentrations for gas sweetening and sulfur recovery operations.

The API data were analyzed to determine the average (weighted) $H_2S$ concentration of the sour gas streams where sulfur recovery is performed and the average (weighted) $H_2S$ concentration of the sour gas streams where only sweetening is performed. This appendix presents: (1) EPA's analysis of the API data, (2) projections on the amount of sulfur in the newly produced sour natural gas for the period of 1983-1987, (3) distribution of the projected sulfur into sweetening and sulfur recovery facilities, and (4) the API Gas Plant Survey Report.

## G.1 ANALYSIS OF THE API DATA

In analyzing API's gas plant data, the gas streams (a total of 731 streams) were distributed into three categories: (1) streams that are naturally sweet and require no further sweetening, (2) streams that are sour and have sweetening without sulfur recovery, and (3) streams that are sour and have sweetening with sulfur recovery. Table G-1 presents a summary of the results of the analysis that was performed on the data. The analysis indicates that the average (weighted) $H_2S$ concentration in the sour gas streams without sulfur recovery is 0.2 mole percent. The 217 streams without sulfur recovery are presented in Table G-2. The analysis also indicates that the average (weighted) $H_2S$ concentration in the sour gas streams where sweetening and sulfur recovery both are performed is 5.8 mole percent. The 61 streams of this type are presented in Table G-3. The analysis as presented in Table G-1 also indicates that about 23.2 mole percent of the present natural gas processing is sour and requires sweetening either with or without sulfur recovery. The remaining 76.8 percent is already sweet and needs no treatment.

## G.2 PROJECTIONS OF NEW SULFUR PRODUCTION IN THE PERIOD OF 1983-1987

Based on the analysis of the API data summarized in Table G-1, the following assumptions were developed to predict the projected sulfur production and recovery from natural gas processing in the period of 1983-1987:

(1) The natural gas that is produced offshore is sweet and new discoveries are expected to be sweet.

(2) An average of 25 percent of the newly discovered onshore natural gas during the projection period of 1983-1987 would be sour, and the remaining 75 percent would be sweet.

(3) An average of about 75.8 percent of the new sour gas would be subjected to sweetening and sulfur recovery. The average $H_2S$ concentration in these sour gas streams would be 5.8 mole percent.

(4) An average of about 24.2 percent of the new sour gas would be subjected to sweetening only. The average $H_2S$ concentration in these sour gas streams would be 0.2 mole percent.

Total estimated new onshore natural gas production for the period was obtained from data supplied by the American Gas Association (Ref. 15, Chapter 9). This projected gas production was used to predict the annual total gas throughput for both sweetening only and sweetening with sulfur recovery operations; utilizing the percentage of total production presented above. The gas throughput and average $H_2S$ concentration in the sour gas stream for sweetening only operations were used to predict the annual and total sulfur in gas streams from sweetening only operations for the 1983-1987 period. The gas throughput and average $H_2S$ concentration in the sour gas stream for sweetening with sulfur recovery operations were used to predict the added new sulfur recovery capacity in the period. Tables G-4 and G-5 present the annual and total sulfur production and recovery figures for the 1983-1987 period.

Table G-4 indicates that the revised fifth year (1987) projection of total new sulfur recovery capacity added to the industry would be 3,244 megagrams per day. Table G-5 indicates that the amount of sulfur removed from the sour gas streams at plants with sweetening only would total 38.9 megagrams per day by the end of 1987.

## G.3 DISTRIBUTION OF THE PROJECTED SULFUR PRODUCTION INTO SWEETENING AND SULFUR RECOVERY FACILITIES

The projection of the fifth year (1983-1987) total sulfur content of sour gas streams with sulfur recovery, 3,244 megagrams per day, was distributed into a number of sulfur recovery facilities of various sulfur intake sizes (sulfur feed rates). The size distribution for the projection period is based on the current size distribution prevailing in the industry. This resulted in 30 new sulfur recovery facilities: 18 facilities of 10 megagrams per day each, 9 facilities of 102 megagrams per day each, 2 facilities of 564 megagrams per day each and 1 facility of 1,016 megagrams per day. Table G-6 indicates the projected population of new sulfur recovery facilities and their sulfur intake size distribution for the 1983-1987 period.

The projected fifth year (1987) total sulfur in sour gas streams with sweetening only, 38.9 megagrams per day, was distributed into sweetening facilities with sulfur intake sizes (rates of sulfur emerging from the sweetening unit) ranging from less than 0.1 to 5.0 megagrams per day. This resulted in 37 facilities. Table G-7 presents the projected population of new sweetening facilities and their sulfur intake size distribution in the 1983-1987 period. The size distribution for the projection period is assumed to remain the same as the current size distribution of 5 or less megagrams per day facilities existing in the industry.

## G.4 API GAS PLANT SURVEY DATA

A copy of the API's original report on their survey of the onshore natural gas plant streams (731 streams) that was submitted in July 1982 to the U.S EPA is presented in Table G-8.

## Table G-1. ANALYSIS OF THE API GAS PLANT SURVEY DATA[a]

## ONSHORE NATURAL GAS PROCESSING

(1) Current total throughput = 36.543 billion cubic feet per day

(2) Throughput with sweetening only = 5.808 billion cubic feet per day
Sweetening only = 15.89% of the total throughput

(3) Throughput with sulfur recovery = 2.685 billion cubic feet per day
Sulfur recovery = 7.35% of the total throughput

(4) Throughput that is already sweet (no treatment needed) = 28.050 billion cubic feet per day

(5) % of the onshore production that is sour =

$$\frac{\text{throughput with sweetening + throughput with sulfur recovery}}{\text{total throughput}}$$

$$= \frac{5.808 + 2.685}{36.543} \times 100$$

$$= \frac{8.493}{36.543} \times 100$$

$$= 23.24\%$$

(6) Well-head sweet gas (no treatment required) = 76.76% of the total throughput

(7) Average (weighted) $H_2S$ in the sour gas throughput with sweetening only
= 0.2 mole percent

(8) Average (weighted) $H_2S$ in the sour gas throughput with sulfur recovery
= 5.8 mole percent

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[a] The API survey was based upon the information published in the OGJ Report presented in Oil and Gas Journal, July 13, 1981, page 94.

### Table G-2. ANALYSIS OF THE API's GAS PLANT SURVEY DATA – SOUR NATURAL GAS STREAMS WITH SWEETENING ONLY[a]

| Plant or stream no. | Current throughput (thousand cubic feet per day) | $H_2S$ mole percent | Sweetening process | Acid gas disposal |
|---|---|---|---|---|
| 4 | 26,823 | 0.02 | DGA | I |
| 8 | 1,700 | 2.75 | MEA | F |
| 9A | 2,700 | 2.7 | MEA | F |
| 37 | 2,300 | 0.01 | IrSp | - |
| 46 | 3,500 | 0.08 | SCIS | - |
| 47 | 800 | 0.6 | SMCS | - |
| 55 | 5,200 | 0.4 | Amine | I |
| 71 | 464 | 0.5 | DEA | I |
| 73B | 400 | 0.2 | Amine | F |
| 89 | 920,000 | 0.02 | - | |
| 116 | 3,500 | 0.31 | DEA | F |
| 125 | 18,000 | 0.0032 | DEA | V |
| 126L | 40,000 | 0.0016 | DEA | V |
| 127 | 100,000 | 0.004 | IrSp | - |
| 128 | 150,000 | 0.004 | DEA | V |
| 129 | 15,000 | 0.004 | DEA | V |
| 130 | 30,000 | 0.004 | DEA | V |
| 131 | 30,000 | 0.004 | DEA | V |
| 133 | 20,000 | 0.004 | - | - |
| 151 | 2,700 | 0.068 | MEA | I |
| 152 | 35,000 | 2.0 | Amine | V |
| 153 | 3,400 | 0.03 | DEA | I |
| 167 | 7,000 | 0.25 | DEA | - |
| 168 | 1,700 | 0.002 | MEA | V |
| 169 | 26,000 | 0.0028 | DEA | V |
| 170 | 65,400 | 0.0018 | DEA | V |

(continued)

Table G-2. Continued

| Plant or stream no. | Current throughput (thousand cubic feet per day) | $H_2S$ mole percent | Sweetening process | Acid gas disposal |
|---|---|---|---|---|
| 197A | 6,000 | 0.016 | IrSp | - |
| 197B | 10,000 | 0.020 | Amine | - |
| 197C | 7,000 | 0.0032 | Sulf | - |
| 196A | 260 | 0.08 | IrSp | - |
| 196B | 1,700 | 0.07 | IrSp | - |
| 196C | 600 | 0.08 | IrSp | - |
| 196F | 500 | 0.15 | Amine | - |
| 196E | 2,500 | 0.0008 | IrSp | - |
| 193 | 13,000 | 0.1 | MEA | F |
| 194 | 1,800 | 8.0 | NSul | F |
| 195A | 350 | 6.7 | Amine | F |
| 198 | 5,000 | 0.158 | DGA | F |
| 200A | 10,000 | 0.825 | Sulf | F |
| 200B | 4,600 | 0.825 | Sulf | F |
| 200C | 10,000 | 0.825 | Sulf | F |
| 200D | 4,600 | 0.825 | Sulf | F |
| 200E | 10,000 | 0.825 | Sulf | F |
| 200F | 4,600 | 0.825 | Sulf | F |
| 200G | 4,600 | 0.825 | Sulf | F |
| 200J | 25,000 | 0.825 | Sulf | F |
| 200K | 4,600 | 0.825 | Sulf | F |
| 200M | 4,600 | 0.825 | Sulf | F |
| 200N | 4,600 | 0.825 | Sulf | F |
| 200P | 4,600 | 0.825 | Sulf | F |
| 200Q | 10,000 | 0.825 | Sulf | F |
| 202 | 2,000 | 0.04 | Amine | V |

(continued)

Table G-2. Continued

| Plant or stream no. | Current throughput (thousand cubic feet per day) | $H_2S$ mole percent | Sweetening process | Acid gas disposal |
|---|---|---|---|---|
| 206 | 35,000 | 0.006 | - | - |
| 207 | 1,000 | 0.003 | IrSp | - |
| 211 | 6,187 | 0.35 | Sulf | F |
| 212 | 4,000 | 2.0 | - | F |
| 213 | 1,028 | 0.0016 | IrSp | - |
| 217 | 3,000 | 0.16 | DEA | I |
| 218 | 18,000 | 0.47 | ADEA | F |
| 223 | 67,430 | 0.24 | MEA | F |
| 225 | 9,590 | 0.03 | MEAG | F |
| 226 | 103,690 | 0.22 | MEAG | F |
| 235 | 9,400 | 0.52 | DGA | F |
| 236 | 16,000 | 2.0 | Amine | F |
| 238 | 3,600 | 1.21 | MEA | I |
| 240 | 8,000 | 0.3 | MEA | F |
| 245 | 19,000 | 0.20 | Amine | F |
| 246 | 61,000 | 0.09 | Amine | F |
| 248 | 65,110 | 0.16 | Amine | F |
| 250 | 20,732 | 0.24 | Amine | F |
| 252 | 4,277 | 0.27 | Amine | F |
| 256 | 4,000 | 0.003 | IrSp | - |
| 257 | 600 | 0.012 | MDEA | F |
| 259 | 6,000 | 1.65 | MEA | I |
| 261 | 900 | 0.0022 | IrSp | - |
| 262 | 1,500 | 0.5 | Amine | F |
| 282 | 6,000 | 0.001 | - | - |
| 293B | 2,000 | 0.44 | - | - |

(continued)

Table G-2. Continued

| Plant or stream no. | Current throughput (thousand cubic feet per day) | $H_2S$ mole percent | Sweetening process | Acid gas disposal |
|---|---|---|---|---|
| 304A | 2,500 | 0.42 | MEA | F |
| 309 | 3,500 | 0.0025 | - | - |
| 315 | 35,000 | 1.58 | MEAA | F |
| 319 | 29,000 | 0.004 | - | - |
| 322 | 69,000 | 0.001 | DEA | I |
| 326 | 24,000 | 0.3 | IrSp | - |
| 336A | 57,000 | 0.046 | - | - |
| 336B | 2,000 | 0.42 | Amine | - |
| 336C | 2,000 | 0.51 | Amine | - |
| 336D | 3,000 | 0.80 | Amine | - |
| 336E | 2,500 | 0.47 | Amine | - |
| 336F | 2,500 | 0.38 | Amine | - |
| 336G | 1,000 | 0.68 | Amine | - |
| 340A | 20,000 | 0.006 | Amine | F |
| 352A | 1,500 | 0.18 | Amine | F |
| 363B | 20,000 | 0.05 | DEA | F |
| 365A | 22,000 | 0.30 | DEA | I |
| 366A | 70,000 | 0.02 | MDEA | I |
| 384 | 40,000 | 0.35 | MEA | I |
| 387 | 12,140 | 0.67 | MEAG | F |
| 388 | 282,490 | N/A | DGA | N/A |
| 389 | 29,820 | N/A | PrC | V |
| 390 | 15,840 | 0.1 | MEAG | F |
| 391 | 124,860 | N/A | PrC | V |
| 392 | 184,680 | 0.06 | DGA | I |
| 393 | 29,370 | 0.41 | DGA | I |

(continued)

Table G-2. Continued

| Plant or stream no. | Current throughput (thousand cubic feet per day) | $H_2S$ mole percent | Sweetening process | Acid gas disposal |
|---|---|---|---|---|
| 394 | 21,910 | 0.05 | MEAG | F |
| 398A | 9,000 | 0.15 | MEA | I |
| 398B | 6,500 | 0.06 | MEA | F |
| 403 | 3,000 | 0.0013 | MEA | F |
| 410 | 10,000 | 0.0034 | IrSp | - |
| 412A | 630 | 0.294 | Amine | F |
| 412B | 26,903 | 0.065 | AmC | V |
| 413 | 8,000 | 0.04 | - | - |
| 414 | 24,834 | 0.04 | AmC | V |
| 420 | 10,000 | 0.015 | IrSp | - |
| 427E | 25,000 | 0.08 | MEA | F |
| 429 | 30,000 | 0.0037 | MDEA | I |
| 430A | 6,000 | 0.0025 | DEA | I |
| 431 | 53,000 | 0.0035 | DEA | V |
| 432 | 103,000 | 0.045 | KCO | I |
| 433B | 6,000 | 0.0015 | - | - |
| 443 | 600 | 0.5 | Amine | I |
| 445 | 140,000 | 0.033 | MEA | F |
| 452A | 80,000 | 0.0011 | DEA | F |
| 452B | 18,200 | 1.0 | DEA | F |
| 452C | 700 | 0.07 | DEA | F |
| 454 | 14,500 | 0.003 | DEA | F |
| 457 | 102,000 | 0.02 | DEA | F |
| 465A | 11,000 | 1.07 | MEA | - |
| 465B | 5,000 | 0.96 | - | - |
| 482 | 24,500 | 0.002 | - | F |

(continued)

Table G-2. Continued

| Plant or stream no. | Current throughput (thousand cubic feet per day) | $H_2S$ mole percent | Sweetening process | Acid gas disposal |
|---|---|---|---|---|
| 484 | 98,000 | 0.0007 | MDEA | V |
| 485 | 6,500 | 1.199 | MEA | I |
| 488 | 8,500 | 5 | IrSp | - |
| 494 | 9,500 | 0.2 | MEA | F |
| 495 | 2,100 | 0.5 | MEA | F |
| 498 | 55,297 | 0.03 | Sulf | I |
| 501 | 24,000 | 0.048 | MEA | F |
| 506 | 5,800 | 0.0016 | DEA | I |
| 507 | 4,000 | 0.02 | DEA | I |
| 508 | 43,000 | 0.06 | DEA | I |
| 511 | 13,000 | 0.004 | - | - |
| 513 | 28,000 | 0.003 | IrSp | - |
| 516 | 9,000 | 0.16 | MEA | F |
| 518 | 27,000 | 0.03 | HPCA | - |
| 520 | 414 | 0.0288 | DEA | F |
| 524 | 1,500 | 3.0 | MEA | F |
| 530 | 10,000 | 0.003 | MEAI | I |
| 543 | 3,700 | 0.2 | MEA | F |
| 553 | 7,800 | 0.024 | DGA | F |
| 554 | 68,700 | 0.5 | DGA | I |
| 555A | 2,800 | 0.0327 | IrSp | - |
| 556 | 7,600 | 0.0093 | IrSp | - |
| 558 | 9,300 | 0.0018 | MEA | F |
| 559B | 2,900 | 0.1179 | MEA | - |
| 559C | 2,900 | 0.0165 | - | - |
| 559D | 2,900 | 0.1198 | - | - |

(continued)

Table G-2. Continued

| Plant or stream no. | Current throughput (thousand cubic feet per day) | $H_2S$ mole percent | Sweetening process | Acid gas disposal |
|---|---|---|---|---|
| 562 | 6,000 | 0.004 | IrSp | - |
| 563 | 6,000 | 0.02 | Amine | F |
| 564 | 3,000 | 0.39 | Amine | F |
| 565 | 8,000 | 1.0 | IrSp | - |
| 567 | 5,000 | 0.8 | IrSp | - |
| 568 | 30,000 | 0.21 | Amine | F |
| 569 | 30,000 | 0.21 | Amine | - |
| 570 | 35,000 | 0.3 | Amine | V |
| 571 | 14,000 | 1.75 | IrSp | - |
| 575 | 65,000 | 0.39 | MEA | F |
| 576 | 90,000 | 0.15 | MEA | F |
| 577 | 10,000 | 0.40 | - | - |
| 578A | 141,000 | 0.03 | MEA | F |
| 578B | 9,000 | 1.0 | DEA | F |
| 579B | 115,000 | 0.07 | - | - |
| 580B | 3,500 | 0.016 | MEA | - |
| 581 | 1,500 | 1.0 | DEA | F |
| 583 | 18,300 | 0.1 | - | I |
| 595 | 1,340 | 2.2 | MEA | I |
| 599 | 8,000 | 0.03 | MEA | F |
| 601 | 2,850 | 0.45 | MEA | F |
| 611 | 35,000 | 0.11 | Amine | F |
| 612 | 25,000 | 0.13 | Amine | F |
| 615 | 6,000 | 0.67 | Amine | F |
| 620 | 185,000 | 0.009 | Amine | F |
| 626 | 16,000 | 0.2 | Amine | F |

(continued)

Table G-2. Continued

| Plant or stream no. | Current throughput (thousand cubic feet per day) | $H_2S$ mole percent | Sweetening process | Acid gas disposal |
|---|---|---|---|---|
| 634 | 9,962 | 0.02 | - | - |
| 635 | 1,795 | 0.06 | - | - |
| 637 | 22,718 | 0.21 | - | - |
| 638A | 500 | 6.0 | Amine | F |
| 641 | 9,468 | 0.3 | - | - |
| 643 | 1,693 | 0.01 | - | - |
| 644 | 699 | 2.44 | - | - |
| 649 | 3,000 | 0.4 | MEA | F |
| 653 | 7,500 | 2.6 | DGA | F |
| 654 | 22,500 | 0.27 | MEA | F |
| 657 | 5,900 | 0.07 | MEA | F |
| 670 | 14,000 | 0.3 | MEA | F |
| 671 | 12,000 | 0.1 | DEA | F |
| 672 | 9,000 | 0.55 | DEA | F |
| 673 | 29,000 | 1.00 | DEA | F |
| 674 | 3,000 | 0.07 | DEA | F |
| 678 | 7,000 | 0.003 | Sulf | V |
| 681 | 10,000 | 0.02 | Sulf | I |
| 684 | 33,000 | 0.55 | - | - |
| 685 | 120,474 | 0.25 | - | - |
| 687 | 5,700 | 0.1 | Amine | F |
| 689 | 8,900 | 0.008 | IrSp | - |
| 690 | 1,900 | 0.008 | - | - |
| 692 | 51,500 | 1.08 | DprC | F |
| 693 | 10,000 | 0.024 | IrSp | - |
| 698 | 15,000 | 0.0032 | Sele | - |

(continued)

Table G-2. Concluded

| Plant or stream no. | Current throughput (thousand cubic feet per day) | $H_2S$ mole percent | Sweetening process | Acid gas disposal |
|---|---|---|---|---|
| 710B | 1,500 | 2.2 | MEA | - |
| 720A | 50,000 | 1.5 | - | - |
| 720B | 5,000 | 0.5 | - | - |
| 720C | 2,000 | 0.5 | DEA | - |
| 724 | 2,500 | 0.35 | DEA | V |
| 729C | 3,000 | 7.0 | DGA | I |
| 730 | 3,000 | 1.6 | MEA | F |

[a] Average (weighted) $H_2S$ concentration in the sour natural gas streams with sweetening only: 0.2 mole percent.

## Table G-3. ANALYSIS OF THE API's GAS PLANT SURVEY DATA - SOUR NATURAL GAS STREAMS WITH SULFUR RECOVERY ONLY[a]

| Plant or stream no. | Current throughput (thousand cubic feet per day) | $H_2S$ mole percent | Sulfur recovery unit | Sulfur recovered, LT/day |
|---|---|---|---|---|
| 1 | 20,000 | 16.75 | Claus 3-stage | 120 |
| 2 | 33,600 | 1.0 | Claus 3-stage | 11.36 |
| 3 | 110,000 | 21.0 | Claus 3-stage | 810 |
| 7 | 12,330 | 3.5 | Claus 3-stage | 12.9 |
| * 17 | 9,000 | 1.5 | Stretford | 5.0 |
| 19 | 1,700 | 8.6 | Claus 2-stage + SCOT | 5.8 |
| 23 | 12,900 | 0.15 | Oxidation/Reduction | 0.77<br>(2.0 design) |
| * 14 | 40,000 | 0.19 | Stretford | 2.5 |
| 67 | 20,800 | 10.1 | Claus 3-stage + SCOT | 78 |
| 68 | 114,900 | 13.5 | Claus 3-stage | 540 |
| * 69 | 21,127 | 0.0416 | Stretford | 0.156 |
| * 70 | 186,850 | 0.064 | Stretford | 1.219 |
| 200H | 30,000 | 1.79 | Claus | 23.1 |
| 204 | 5,606 | 7.6 | Claus | 13 |
| 205 | 98,114 | 36.3 | Claus | 1,089 |
| 221 | 88,000 | 1.096 | Claus 2-stage | 27.04 |
| 224 | 190,270 | 0.23 | Claus 3-stage | 9.62 |
| 227 | 29,975 | 2.23 | Claus 3-stage | 23.9 |
| 228 | 100,000 | 0.5 | Claus 3-stage | 18 |
| 229 | 34,982 | 2.7 | Claus 3-stage | 26 |
| 231 | 28,000 | 1.375 | Claus 3-stage | 18 |
| 237 | 30,500 | 1.1 | Claus 3-stage | 11.46 |
| 249 | 54,000 | 0.83 | Claus | 16.1 |
| 253 | 38,400 | 2.5 | Claus 3-stage | 34.1 |
| 254 | 140,000 | 0.3 | Claus | 10 |
| 258 | 53,700 | 3.85 | Claus | 82.9 |
| 260 | 16,780 | 9.5 | CBA | 58.8 |
| 361 | 125,000 | 0.25 | Claus | 12.1 |

(continued)

Table G-3. Continued

| Plant or stream no. | Current throughput (thousand cubic feet per day) | $H_2S$ mole percent | Sulfur recovery unit | Sulfur recovered, LT/day |
|---|---|---|---|---|
| 364A | 69,000 | 0.46 | Claus 3-stage | 11.3 |
| 365B | 12,000 | 11.00 | Claus 3-stage | 47 |
| 365C | 41,000 | 4.42 | Claus 3-stage | 65 |
| 385 | 107,360 | 1.03 | Claus 3 stage | 31.13 |
| 386 | 15,230 | 6.29 | Claus 3-stage | 35.59 |
| 402 | 20,000 | 5.0 | Claus 2-stage | 24 |
| 428 | 6,000 | 1.7 | Claus 2-stage | 3.5 |
| 476 | 23,500 | 2.04 | Claus 1-stage | 10.47 |
| 478 | 45,000 | 4.209 | Claus 3-stage | 63 |
| 492 | 2,000 | 4.8 | Claus 2-stage | 2 |
| 505 | 9,400 | 3.8 | Claus | 12.35 |
| 537 | 17,000 | 3.0 | Claus 2-stage | 15 |
| 546 | 59,000 | 1.28 | Claus 2-stage | 21.5 |
| 557 | 1,000 | 5.25 | Claus 3-stage | 3.10 |
| 566 | 4,000 | 3.6 | Claus 3-stage | 4.5 |
| 585 | 6,700 | 30.00 | Claus 3-stage | 76.0 |
| 596 | 30,650 | 2.0 | Claus 2-stage | 7.56 |
| 616 | 28,512 | 6.46 | Claus 3-stage | 113.3 |
| 619 | 7,778 | 27.20 | Claus 2-stage | 68.4 |
| 630 | 1,800 | 7.50 | Claus | 4.5 |
| 633 | 37,487 | 2.15 | Claus | 29.0 |
| 640 | 78,666 | 1.05 | Claus | 27.0 |
| 647 | 62,063 | 1.8 | Claus | 37.0 |
| 651 | 60,000 | 7.7 | Claus 3-stage | 174 |
| 656 | 10,400 | 3.8 | Claus 2-stage +CBA | 14.4 |
| 658 | 19,900 | 4.7 | Claus 2-stage | 24.21 |
| 660 | 34,300 | 3.5 | Claus 2-stage | 27.6 |
| 661 | 10,150 | 0.09 | Claus 2-stage | 3.1 |
| 663 | 11,100 | 5.4 | Claus 2-stage | 26.0 |
| 675 | 70,352 | 5.52 | Claus | 118 |

(continued)

Table G-3. Concluded

| Plant or stream no. | Current throughput (thousand cubic feet per day) | $H_2S$ mole percent | Sulfur recovery unit | Sulfur recovered, LT/day |
|---|---|---|---|---|
| 677 | 38,400 | 31.85 | Claus + SCOT | 384 |
| 682 | 20,528 | 1.16 | Claus | 7 |
| 701 | 50,000 | 3.8 | Claus 2-stage + ATS | 68 |
| 713 | 3,200 | 14.9 | Claus 3-stage | 11.2 |
| 726 | 5,400 | 6.0 | Claus | 14.5 |
| 727 | 11,700 | 10.0 | Claus | 50 |
| 729B | 8,000 | 12.0 | Claus + CBA | 36.1 |

[a] Average (weighted) $H_2S$ concentration in the sour natural gas streams with sulfur recovery only: 5.8 mole percent.

*Considered offshore stream.

Table G-4. PROJECTED NEW SULFUR RECOVERY CAPACITIES FOR THE PERIOD OF 1983-1987

| Year | New Onshore natural gas production in the specified year $10^6$ ft$^3$ | Onshore production that is sour[a] $10^6$ ft$^3$ | Throughput for sweetening only[b] $10^6$ ft$^3$ | Throughput for sulfur recovery[c] $10^6$ ft$^3$ | Average $H_2S$ in the sour natural gas streams processed for sulfur recovery, mole percent | New sulfur recovery capacity added in the specified year $10^3$ Mg/y | New sulfur recovery capacity added in the specified year[d], Mg/d |
|---|---|---|---|---|---|---|---|
| 1983 | 565,000 | 141,250 | 36,380 | 104,870 | 5.8 | 232.2 | 663.4 |
| 1984 | 593,000 | 148,250 | 38,180 | 110,070 | 5.8 | 244.1 | 697.4 |
| 1985 | 616,000 | 154,000 | 39,670 | 114,330 | 5.8 | 253.3 | 723.7 |
| 1986 | 483,000 | 120,750 | 31,100 | 89,650 | 5.8 | 199.2 | 569.1 |
| 1987 | 501,000 | 125,250 | 32,260 | 92,990 | 5.8 | 206.5 | 590.0 |
| Total in period of 1983-1987 | | | | | | | |
| | 2,758,000 | 689,500 | 177,590 | 511,910 | | 1,135 | 3,244 |

[a] Derived values based upon an average of 25 percent of the onshore production in the period of 1983-1987 being sour and the remaining 75 percent being sweet.

[b] Derived values based upon 25.7 percent of the sour onshore new discovery for the projected period 1983-1987 processed for sweetening only.

[c] Derived values based upon 74.3 percent of the sour onshore new discovery for the projected period 1983-1987 processed for sulfur recovery.

[d] 350 days/year operating schedule.

## Table G-5. SULFUR IN SOUR NATURAL GAS STREAMS WITH SWEETENING ONLY IN THE PERIOD 1983-1987

| Year | Throughput for sweetening only $10^6$ ft$^3$ | Average $H_2S$ in the sour natural gas stream, mole percent | Sulfur in the gas streams added in the specified year[a] $10^3$ Mg/y | Mg/d |
|---|---|---|---|---|
| 1983 | 36,380 | 0.2 | 2.76 | 7.96 |
| 1984 | 38,180 | 0.2 | 2.91 | 8.34 |
| 1985 | 39,670 | 0.2 | 3.02 | 8.68 |
| 1986 | 31,100 | 0.2 | 2.38 | 6.79 |
| 1987 | 32,260 | 0.2 | 2.49 | 7.06 |
| Total in the period of 1983-1987 | | | | |
| | 177,590 | | 13.6 | 38.9 |

[a] 350 days/year operating schedule.

Table G-6. PROJECTED POPULATION OF NEW SULFUR RECOVERY FACILITIES AND THEIR SULFUR INTAKE SIZE DISTRIBUTION, 1983-1987

| Year | Average size (Mg/d) | | | | Total |
|---|---|---|---|---|---|
| | 10 | 102 | 564 | 1,016 | |
| 1983 | 3 | 1 | - | - | 4 |
| 1984 | 4 | 2 | 1 | - | 7 |
| 1985 | 4 | 3 | - | 1 | 8 |
| 1986 | 4 | 2 | - | - | 6 |
| 1987 | 3 | 1 | 1 | - | 5 |
| Number | 18 | 9 | 2 | 1 | 30 |
| Percent of total number | 60 | 30 | 6.7 | 3.3 | 100 |
| New sulfur recovery capacity, Mg/d | 180 | 918 | 1,128 | 1,016 | 3,244 |

Table G-7. PROJECTED POPULATION OF NEW SWEETENING FACILITIES AND THEIR SULFUR INTAKE SIZE DISTRIBUTION, 1983-1987

| Sulfur intake size (avg.), Mg/d | Percent of total sulfur, % | Amount of sulfur in the specified size, Mg/d | Number of the facilities |
|---|---|---|---|
| <0.1 (0.034 avg.) | 0.95 | 0.369 | 11 |
| 0.1 | 0.2 | 0.078 | 1 |
| 0.2 | 1.3 | 0.506 | 3 |
| 0.3 | 1.05 | 0.408 | 1 |
| 0.4 | 1.8 | 0.700 | 2 |
| 0.5 | 1.75 | 0.681 | 1 |
| 0.6 | 1.8 | 0.700 | 1 |
| 0.7 | 1.05 | 0.409 | 1 |
| 0.8 | 0.8 | 0.311 | 1 |
| 0.9 | 3.15 | 1.23 | 1 |
| 1.0 | 13.15 | 5.12 | 5 |
| 2.0 | 16.05 | 6.24 | 3 |
| 3.0 | 26.75 | 10.41 | 3 |
| 4.0 | 24.95 | 9.71 | 2 |
| 5.0 | 5.25 | 2.04 | 1 |
| TOTAL | 100% | 38.9 | 37 |

## Table G-8. THE API GAS PLANT SURVEY REPORT

July 15, 1982

### ABBREVIATIONS
### API GAS PLANT SURVEY REPORT

Gas Sweetening Process (Columns 8 & 12)
- AB -- Absorption
- ADEA -- Absorption DEA
- AIrSp -- Amine & Iron Sponge
- AmC -- Amine $CO_2$
- Bfd -- Benfield
- DEA -- DEA
- DEAC -- DEA $CO_2$
- DGA -- DGA
- DPrC -- DEA Propylene Carbonate
- ETMR -- Ethane Treating, MEA Regeneration
- HP -- Hot Pot
- HPCA -- Hot Potassium Carbonate and Amine
- IrCl -- Iron Chloride
- IrSp -- Iron Sponge
- KCO -- $K_2CO_2$
- MDEA -- MDEA
- MDGA -- MDGA
- MEA -- MEA
- MEAA -- MEA with Alumina Selicate Bed
- MEAG -- MEA/Glycol
- MEAI -- MEA & Iron Sponge
- NSul -- NaHS & Sulfinol
- PrC -- Propylene Carbonate
- Sele -- Selexol
- SCIS -- Static Mixer Caustic Scrub, Iron Sponge
- SlSw -- Slurry Sweetening
- SMCS -- Static Mixer Caustic Scrub
- Stret -- Stretford
- Sulf -- Sulfinol
- Sul2 -- 2 Sulfinol Units

Acid Gas Disposition (Column 13)
- F -- Flare
- I -- Incinerate
- V -- Vent
- SR -- Sulfur Recovery

Type Sulfur Recovery Plant (Column 14)
- 1StCl -- 1 Stage Claus
- 2StCl -- 2 Stage Claus
- 3StCl -- 3 Stage Claus
- 23Cl -- Two 3 Stage Claus Units
- ClCBA -- Claus & CBA
- MDGA -- MDGA
- OR -- Oxidation Reduction
- Stret -- Stretford
- WCFB -- Wood Chip Fixed Bed

Type Tail Gas Treatment (Columns 19 & 20)
- ATS -- ATS Plant
- Incin -- Incineration
- $SO_2$ -- $SO_2$ at 4.4 MOLE/Hr

Sulfur Disposition (Column 21)
- CRC -- Chemical Disposal (Land)
- LdDl -- Land Disposal
- LaFl -- Land Fill
- SalTr -- Sales/Truck
- SBP -- Sulfur Owned By Product
- TrLiq -- Truck Liquid

# API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | Total Plant System | | Field Gas Sweeting | | | | | Sweet-ening Proc-ess | Acid Gas Disp | Sulfur Recovery Plant | | | | | Tail Gas Treating | | |
| | | | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Proc-ess Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After | | | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 1 | 31000 | 20000 | 16.75 | 5.4 | No | --- | --- | --- | --- | Sulf | SR<br>(Also I250 and F250) | 3StCl | 182 | 125 | 120 | 11200 | Incin | 11200 | Sales |
| 2 | 57000 | 33600 | 1.0 | 3.75 | No | | --- | --- | --- | Bfd | SR | 3StCl | 22 | 13.37 | 11.36 | 4500 | --- | --- | Sales |
| 3 | 40000 | 32000 | --- 21.0 | 0.2 | Yes | Sul2 | 110000 | 21 | --- | Sulf | SR | 23SCL | 880 | 870 | 810 | 134400 | --- | --- | SalTr |
| 4 | 50000 | 26823 | 0.02 | 5.54 | No | --- | --- | --- | --- | DGA | I | 2StCl | 12 (not in oper. due to low $H_2S$ cont. of gas) | | | | | | |
| 5 | 32000 | 23000 | -0- | 0.09 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 6 | 19500 | 19500 | -0- | 0.12 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 7 | 50000 | 12330 | 3.50 | 1.6 | No | --- | --- | --- | --- | MEA | SR<br>(Also V100) | 3StCl | 24 | 13.4 | 12.9 | 1120 | 1125 | 1120 | --- |
| 8 | 3000 | 1700 | 2.75 | 5.0 | No | --- | --- | --- | --- | MEA | F100 | --- | --- | --- | --- | --- | --- | --- | --- |
| 9 | 15000 | | | | | | | | | | | | | | | | | | |
| 9A | | 2700 | 2.7 | 5.8 | No | --- | --- | --- | --- | MEA | F116 | --- | --- | --- | --- | --- | --- | --- | --- |
| 9B | | 9300 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 10 | 25000 | 23600 | -0- | 1.55 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 11 | 25000 | 21000 | 0.0008 | 0.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
|  |  |  | Total Plant System |  | Field Gas Sweeting |  |  |  |  |  |  | Sulfur Recovery Plant |  |  |  |  | Tail Gas Treating |  |  |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After | Sweetening Process | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 12 | 80000 | 65000 | -0- | 1.2 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 13 | 14000 | 5700 | Nil | Nil | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| → 14 | 60000 | --- | --- | --- | --- | --- | --- | --- | --- | --- | SR | Stret | 2.8 | --- | --- | 672 | None | --- | LdDl |
| 14O | (Offshore) | 12000 | 0.0037 | 3.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 14L | (Onshore) | 5000 | 0.12 | 5.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 15 | 190000 | 24700 | 0.0004- | 1.51 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 16 | 3000 | 7000 | 0.0004- | 19.00 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 17 | 20000 | 9000 | 1.5 | 12.0 | No | --- | --- | --- | --- | Stret | SR | Stret | 11 | 5 | 5 | -0- | --- | --- | --- |
| 18 | 70000 | 27700 | Nil | 0.35 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| * 19 | --- | --- | --- | --- | Yes | MDGA | 1700 | 8.6 | --- | MDGA | SR | 2StCl | 10 | 5.8 | 5.8 | 12.1 | Scott | 12.1 | Sales |
| (Field sweetening site. MDGA capacity = 3700 MCFD. Inlet MOL % $CO_2$ = 19.0) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 20 | 50000 | 19000 | -0- | 6.86 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 21 | 50000 | 14000 | -0- | 18.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 22 | 46000 | 63000 | 0.003 | 0.47 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 23 | 35000 | 12900 | 0.15 | 9.1 | No | --- | --- | --- | --- | IrCl | SR | OR | 2.0 | 0.77 | 0.77 | -0- | --- | -0- | CDL |
| (Note Re Col. 20: Recovered as Sulfur, FES, & $H_2S$ slurry 220 BBL/Month.) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 24 | 20000 | 14000 | -0- | 1.1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 25 | 3000 | 3130 | -0- | 1.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 26 | 20000 | 6660 | -0- | 6.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 27 | 15000 | 7000 | 0.0032- | 6.94 | Yes | Stret | 4500 | 0.144 | 0.0024 | --- | --- | Stret | 0 .74 | (not in service) |  |  | --- | --- | --- |
| ($H_2S$ in Col. 7 exceeds Col. 14 because 2500 MCFD of total plant throughput is not sweetened.) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 28 | 3500 | 1570 | Trace | 26 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 29 | 16000 | 3200 | -0- | 0.9 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 30 | 18000 | 1500 | -0- | 0.4 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 31 | 18000 | 7500 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 31A | 500 | --- | Trace | 0.7 | Yes | IrSp | --- | Trace | -0- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 31B | 8500 | --- | --- | 0.1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 31C | 9000 | --- | --- | 0.6 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 32 | 12500 | 3500 | -0- | 2.2 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 33 | 45000 | 1238 | --- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 34 | 13000 | 6810 | --- | 1.93 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
|  |  |  |  |  |  |  |  |  |  |  |  | Sulfur Recovery Plant |  |  |  |  |  |  |  |
|  |  |  | Total Plant System |  | Field Gas Sweeting |  |  |  |  |  |  |  |  |  |  |  | Tail Gas Treating |  |  |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Proc-ess Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After | Sweet-ening Proc-ess | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 35 | 26600 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 350A | (Offshore) | 2500 | -0- | 2.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 350B | (Offshore) | 8200 | 0.0004 | 2.8 | Yes | Stret | 10000 | 0.02 | 0.0002 | --- | SR | Stret | 2.824 | 0.0753 | 0.0746 | 1.68 | --- | --- | --- |
| (Notes: a. Re Col 9. 8200 MCFD from stretford unit to Plant No. 35.<br>b. Re Col 21. Due to a low $H_2S$ concentration, no sulfur cake produced. Liquor containing sulfur periodically barged off and disposed of as hazardous waste.) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 36 | 30000 | 60 | 0.0001 | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 37 | 3500 | 2300 | -0- | 9.0 | Yes | IrSp | 2000 | 0.01 | -0- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 38 | 15000 | 10000 | -0- | 4.7 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 39 | 10000 | 5900 | 0.00004 | 11.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 40 | 60000 | 56700 | 0.00003 | 3.8 | No | --- | --- | --- | --- | PrC | V33 | --- | --- | --- | --- | --- | --- | --- | --- |
| 41 | 5000 | 1500 | 0.0005 | 4.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 42 | 45000 | 16000 | 0.00004 | 10.0 | No | --- | --- | --- | --- | HP | V20 | --- | --- | --- | --- | --- | --- | --- | --- |
|  |  |  |  |  |  |  |  |  |  | (Liquid sales $CO_2$ product. Used only when not liquifying.) |  |  |  |  |  |  |  |  |  |
| 43 | 7000 | 600 | -0- | 0.4 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 44 | 110000 | 11600 | 0.00003 | 0.82 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 45 | 3000 | 1000 | -0- | 7.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 46 | --- | --- | --- | --- | Yes | SCIS | 3500 | 0.08 | -0- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| (Non-producing field sweeting site. SCIS capacity = 6000 MCFD. Inlet MOL % $CO_2$ = 15.0) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 47 | --- | --- | --- | --- | Yes | SMCS | 800 | 0.6 | 0.08 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| (Non-producing field sweetening site. SMCS capacity = 1000 MCFD. Inlet MOL % $Co_2$ = 4.8) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | | | | | | | | Sulfur Recovery Plant | | | | | | | |
| | | | Total Plant System | | Field Gas Sweeting | | | | | | | | | | | | Tail Gas Treating | | |
| | | | | | | | | MOL % $H_2S$ | | | | | | | | | | | |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Proc-ess Used | Vol (MCFD) | Before | After | Sweet-ening Proc-ess | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 48 | 10000 | 6500 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 49 | 2000 | 1700 | -0- | 1.65 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 50 | 350000 | 240000 | -0- | 1.68 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 51 | 20000 | 14000 | -0- | 0.25 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 52 | 20000 | 6100 | Nil | 2.60 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 53 | 22500 | 10700 | Nil | 2.20 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 54 | 28000 | 1500 | 0.15 | 1.30 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 55 | 10000 | 5200 | 0.40 | 2.7 | Yes | Cau | 5200 | 0.40 | 0.0004- | IrSp<br>Amine | SR<br>I | WCFB<br>--- | 0.251<br>--- | 0.0895<br>--- | 0.0893<br>--- | 0.448<br>--- | ---<br>--- | ---<br>--- | LaFl<br>--- |
| (Caustic sweetening & Amine unit about to start-up. Capacity = 10,000 MCFD) | | | | | | | | | | | | | | | | | | | |
| 56 | 1000 | 230 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 57 | 5500 | 5000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 58 | 25000 | 16000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 59 | 30000 | 15900 | --- | 3.48 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 60 | 10000 | 8000 | --- | 2.48 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 61 | 53000 | 40800 | --- | 1.13 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 62 | 150000 | 125300 | --- | 2.53 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 63 | 30000 | 20238 | -0- | 2.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 64 | 1500 | 6913 | -0- | 1.75 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 65 | 7500 | 3500 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | Total Plant System | | Field Gas Sweeting | | | | | | | Sulfur Recovery Plant | | | | | Tail Gas Treating | | |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Proc-ess Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After | Sweet-ening Proc-ess | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf Equiv #/Day | Sulf. Disp. |
| 66 | 750000 | 700000 | -0- | 0.60 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 67 | 9200 | 9100 | --- | 0.2 | Yes | Sulf | 20800 | 10.1 | --- | Sulf | SR | 3StCl | 80 | 79 | 78 | 1340 | Scott | 900 | SalTr |
| 68 | 90000 | --- | --- | 0.01 | Yes | --- | 114900 | 12-15 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 68A | | 37500 | --- | --- | --- | DEA | --- | --- | --- | DEA | SR | 3StCL | 240 | 203 | 195 | 17920 | --- | --- | SalTr |
| 68B | | 13100 | --- | --- | --- | DEA | --- | --- | --- | DEA | SR | 3StCl | 80 | 82 | 79 | 6720 | --- | --- | SalTr |
| 68C | | 19300 | --- | --- | --- | Sulf | --- | --- | --- | Sulf | SR | 3StCl | 150 | 134 | 129 | 11200 | --- | --- | SalTr |
| 68D | | 12600 | --- | --- | --- | Bfd | --- | --- | --- | Bfd | SR | 2StCl | 136 | 151 | 137 | 31360 | --- | --- | SalTr |
| 69 | 65000 | 20127 | 0.0416 | 0.555 | Yes | Stret | 20127 | 0.0416 | 0.0004 | Stret | SR | Stret | 1 | 0.315 | 0.156 | 356 | --- | --- | Sales |
| 70 | 650000 | 186850 | 0.064 | 1.000 | Yes | Sulf | 186850 | 0.064 | 0.0004 | Sulf | SR | Stret | 12 | 4.496 | 1.219 | 7341 | --- | --- | Sales |
| 71 | 1500 | 464 | 0.50 | 0.50 | No | --- | --- | --- | --- | DEA | I | --- | --- | --- | --- | --- | --- | --- | --- |

(Note 3: Sulfur recovery plant only operates during withdrawal season (November through March).)

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## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
|  |  |  | Total Plant System |  | Field Gas Sweeting |  |  |  |  |  |  | Sulfur Recovery Plant |  |  |  |  |  |  |  |
|  |  |  |  |  |  |  |  | MOL % $H_2S$ |  |  |  |  |  |  |  |  | Tail Gas Treating |  |  |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | Before | After | Sweetening Process | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 72 | 240000 | 74540 | -0- | 0.04 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 73 | 16000 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 73A |  | 10000 | -0- | 0.2 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 73B |  | 400 | 0.2 | 1.4 | No | --- | --- | --- | --- | Amine | F30 | --- | --- | --- | --- | --- | --- | --- | --- |
| 74 | 10000 | 10000 | -0- | 0.2 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 75 | 20000 | 20000 | -0- | 0.2 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 76 | 3000 | 3000 | -0- | 0.1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 77 | 1000000 | 730000 | -0- | 0.67 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 78 | 210000 | 103200 | -0- | 0.03 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 79 | 76000 | 25000 | Trace | 0.30 | No | --- | --- | --- | --- | MEA | V91 | --- | --- | --- | --- | --- | --- | --- | --- |
| 80 | 100000 | 72000 | -0- | 0.14 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 81 | 550000 | 352000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 82 | 30000 | 22000 | -0- | 0.14 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 83 | 65000 | 31000 | -0- | 0.15 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 84 | 164000 | 133000 | -0- | 0.10 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 85 | 89000 | 66000 | -0- | 0.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 86 | 5600 | 3700 | -0- | 0.27 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 87 | 30000 | 8500 | Nil | 0.16 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 88 | 25000 | 12500 | Nil | 0.23 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 89 | 940000 | 920000 | 0.02 | 0.05 | Yes | (By others to 0.0004 MOL% $H_2S$.) |  |  |  |  | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 90 | 620000 | 150300 | --- | 0.13 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 91 | 200000 | 175000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 92 | 845000 | 590000 | 0.00011 | 1.06 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

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API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | | | | | | | | Sulfur Recovery Plant | | | | | | | |
| | | | Total Plant System | | Field Gas Sweeting | | | | | | | | | | | | Tail Gas Treating | | |
| Plant or Stream No. | Current Capacity (MCFD) | Current Throughput (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After | Sweetening Process | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 83 | [illegible] | 4901 | -0- | 0.01 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 84 | 300000 | 210000 | -0- | 0.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 85 | 75000 | 2500 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 86 | 50000 | 25000a | Nil | 3.1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 87 | 80000 | 45000a | Nil | 3.2 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 88 | 40000 | 31000a | -0- | 0.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 89 | 90000 | 35000a | -0- | 0.75 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 100 | --- | --- | --- | --- | Yes | Amine | 50000 | Nil | Nil | --- | --- | V40 | --- | --- | --- | --- | --- | --- | --- |

($CO_2$ sweetening site. Gas not processed downstream. Throughput based on 6 months average. Amine plant capacity = 120000. Inlet MOL % $CO_2$ = 7.0)

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| 101 | 67000 | 55000 | -0- | 2.2 | No | --- | --- | --- | --- | MEA | V41<br>(Vent through boiler stack.) | --- | --- | --- | --- | --- | --- | --- | --- |
| 102 | 125000 | 75000 | -0- | 0.6 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 103 | 567000 | 120000 | -0- | 7.78 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 104 | 300000 | 18613 | -0- | 1.19 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 105 | 130000 | 41403 | -0- | 1.7 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 106 | 400000 | 14500 | -0- | 0.6 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 107 | 85000 | 55000 | 0.0004- | 0.81 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 108 | 300000 | 42000 | Nil | 0.90 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 109 | 13000 | 20200 | Nil | 0.60 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 110 | 70000 | 34000 | Nil | 1.20 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 111 | 30000 | 17800 | Nil | 1.10 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 112 | 250000 | 100000 | Nil | 2.10 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 113 | 200000 | 28100 | Nil | 0.60 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 114 | 1300000 | --- | -0- | 0.508 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | | 8300 | | | | | | | | | | | | | | | | | |
| [illegible] | | 76700 | | | | | | | | | | | | | | | | | |
| 115 | 65000 | 15000 | -0- | 1.021 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 116 | --- | --- | --- | --- | Yes | DEA | 3500 | 0.31 | --- | DEA | F150 | --- | --- | --- | --- | --- | --- | --- | --- |

(Sweetening site. Capacity = 30000. Inlet MOL % $CO_2$ = 5.5)

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
|  |  |  | Total Plant System |  | Field Gas Sweeting |  |  |  |  |  |  | Sulfur Recovery Plant |  |  |  |  | Tail Gas Treating |  |  |
| Plant or Stream No. | Current Capacity (MCFD) | Current Throughput (MCFD) | H2S (MOL%) | CO2 (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % H2S Before | MOL % H2S After | Sweetening Process | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 117 | 470000 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 117LA (Onshore A) | 20000 |  | -0- | 1.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 117LB (Onshore B) | 6310 |  | -0- | 0.1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 117OA (Offshore A) | 69600 |  | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 117OB (Offshore B) | 9117 |  | -0- | 1.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 117OC (Offshore C) | 226481 |  | -0- | 0.1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 117OD (Offshore D) | 8705 |  | -0- | 0.6 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 118 | 950000 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 118AL (A onshore) | 25277 |  | -0- | 1.4 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 118AO (A Offshore) | 237166 |  | -0- | 1.1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 118BL (B onshore) | 6217 |  | -0- | 1.6 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 118BO (B Offshore) | 275540 |  | -0- | 1.6 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 119 | 125000 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| (Plant processes only offshore gas.) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 119A (Offshore A) | 12375 |  | -0- | 0.9 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 119B (Offshore B) | 20930 |  | -0- | 0.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 1200 | 500000 | 415196 | -0- | 0.2 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| (Plant processes only offshore gas.) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 1210 | 110000 | 153000 | -0- | 0.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| (Plant processes only offshore gas.) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 1220 | 125000 | 82495 | -0- | 0.6 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| (Plant processes only offshore gas.) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 123 | 80000 | 25300 | 0.001 | 0.40 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 124 | 50000 | 58000 | 0.0016- | 1.00 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 125 | 20000 | 14000 | 0.0032- | 3.00 | No | --- | --- | --- | --- | DEA | V70 | --- | --- | --- | --- | --- | --- | --- | --- |
| 126 | 300000 | --- | --- | --- | --- | --- | --- | --- | --- | DEA | V40 | --- | --- | --- | --- | --- | --- | --- | --- |
| 1260 (Offshore) |  | 180000 | 0.0016- | 2.20 | No |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 126L (Onshore) |  | 66000 | 0.0016- | 2.20 | No |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 127 | 210000 | 100000 | 0.004 | 8 | Yes | IrSp | 100000 | 0.004 | 0.001 | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | Total Plant System | | Field Gas Sweetening | | | | | | | Sulfur Recovery Plant | | | | | Tail Gas Treating | | |
| Plant or Stream No. | Current Capacity (MCFD) | Current Throughput (MCFD) | H₂S (MOL %) | CO₂ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % H₂S Before | MOL % H₂S After | Sweetening Process | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| [illegible] | 150000 | Startup | 0.004 | 8 | Yes | DEA | 150000 | 0.004 | -0- | DEA | V50 | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 15000 | Startup | 0.004 | 8 | Yes | DEA | 15000 | 0.004 | -0- | DEA | V50 | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 50000 | Startup | 0.004 | 8 | Yes | DEA | 30000 | 0.004 | -0- | DEA | V50 | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 30000 | Startup | 0.004 | 8 | Yes | DEA | 30000 | 0.004 | -0- | DEA | V77 | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 190000 | 90000 | -0- | 1.29 (Chromatograph) | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 25000 | 20000 | 0.004 | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 15000 | 3000 | -0- | 1.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 10000 | 2000 | -0- | 1.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 4000 | 3000 | -0- | 1.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 200000 | 170000 | -0- | 0.106 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 100000 | 25000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 950000 | --- | --- | 0.7 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | (Unknown) | 515000 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | (Unknown) | 200000 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 900000 | 190000 | --- | 1.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 100000 | --- | --- | 0.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | (Unknown) | 50000 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 200000 | 75000 | --- | 1.6 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 110000 | 61000 | --- | 1.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 65000 | 21000 | --- | 1.4 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 27000 | 27000 | -0- | 0.7 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 800000 | 145000 | -0- | 0.6 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 30000 | 7500 | -0- | 0.7 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 875000 | 512000 | -0- | 0.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 800000 | 350000 | -0- | 1.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 900000 | 900000 | -0- | 0.3 | No | 83 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 12000 | 11000 | 0.0004 | 1.1 | Yes | 83 | 2700 | 0.048 | 0.0004 | MEA | 140 | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | --- | --- | --- | --- | Yes | Amine | 35000 | 2.0 | Nil | Amine | V65 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = [illegible] MCFD. Sales MOL % CO₂ = 6.69) | | | | | | | | | | | | | | | | | | | |
| [illegible] | 7500 | 3400 | 0.03 | 2.27 | No | --- | --- | --- | --- | DEA | 130 | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 65000 | 15360 | -0- | 0.51 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | [illegible] | [illegible] | -0- | 0.44 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

# API GAS PLANT SURVEY

| 1<br>Plant or Stream No. | 2<br>Current Capacity (MCFD) | 3<br>Current Through-put (MCFD) | 4<br>Total Plant Intake<br>$H_2S$ (MOL%) | 5<br><br>$CO_2$ (MOL %) | 6/7<br>Field Gas Sweetening<br>Yes or No | 8<br><br>Process Used | 9<br><br>Vol (MCFD) | 10<br><br>MOL % $H_2S$ Before | 11<br><br>After | 12<br>Sweet-ening Process | 13<br>Acid Gas Disp | 14<br>Sulfur Recovery Plant<br>Type | 15<br><br>Design Cap LT/Day | 16<br><br>(A) Avail. Sulf LT/Day | 17<br><br>(B) Corres Meas. Rec. LT/Day | 18<br><br>(A)-(B) Equiv. Sulf. Diff. #/Day | 19<br>Tail Gas Treating<br>Type | 20<br><br>Exit Sulf. Equiv #/Day | 21<br><br>Sulf. Disp. |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| [illegible] | [illegible] | [illegible] | | | | | | | | | | | | | | | | | |
| [illegible] | | | --- | 0.46 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | | | --- | 0.57 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | | | --- | 0.35 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | | | --- | 0.72 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | | | --- | 1.30 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | | | --- | 0.39 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 200000 | --- | --- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 140000 | 130000 | --- | 1.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 750000 | 620000 | --- | 0.007 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 250000 | 220000 | --- | 0.007 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] processes only ethane [illegible] | | | | | | | | | | | | | | | | | | | |
| [illegible] | 40000 | 1500 | Nil | 0.55 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 200000 | 20015 | -0- | 1.98 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 83000 | 22583 | -0- | 0.60 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 80000 | 15027 | -0- | 0.28 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 93000 | 15648 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 120000 | 25000 | -0- | 2.17 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 814830 | 409618 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 80000 | 19300 | --- | 1.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | --- | --- | --- | --- | Yes | DEA | 970 | 0.25 | --- | DEA | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] plant. Capacity = 7000 MCFD. Inlet MOL % $CO_2$ = 2.75) | | | | | | | | | | | | | | | | | | | |
| [illegible] | --- | --- | --- | --- | Yes | DEA | 1700 | 0.002 | --- | MEA | 965 | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] plant. Capacity = 30000 MCFD. Inlet MOL % $CO_2$ = 9.0) | | | | | | | | | | | | | | | | | | | |
| [illegible] | --- | --- | --- | --- | Yes | DEA | 20000 | 0.0028 | --- | --- | 965 | --- | --- | --- | --- | --- | --- | 3110 | --- |
| [illegible] plant. Capacity = 70000 MCFD. Inlet MOL % $CO_2$ = 6.0) | | | | | | | | | | | | | | | | | | | |
| [illegible] | --- | --- | --- | --- | Yes | DEA | 65600 | 0.0012 | --- | --- | 965 | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] plant. Capacity = 90000 MCFD. Inlet MOL % $CO_2$ = 8.5) | | | | | | | | | | | | | | | | | | | |
| [illegible] | 25000 | 6400 | --- | 0.2 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| [illegible] | 80000 | 10600 | --- | 1.6 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | | | | | | | | Sulfur Recovery Plant | | | | | | | |
| | | | Total Plant System | | Field Gas Sweeting | | | | | | | | | | | | Tail Gas Treating | | |
| Plant or Stream No. | Current Capacity (MCFD) | Current Throughput (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % $H_2S$ Before | After | Sweetening Process | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 173 | 120000 | 31200 | --- | 2.1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 174 | 95000 | 27300 | --- | 2.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 175 | 297000 | 86200 | --- | 1.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 176 | 71000 | 4341 | Nil | 0.31 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 177 | 100000 | Startup | --- | 1.00 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 178 | 12000 | 12896 | --- | 0.19 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 179 | 1200000 | 979505 | --- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 180 | 23000 | 6628 | --- | 0.37 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 181 | 120000 | 78306 | --- | 1.54 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 182 | 120000 | 6767 | Nil | 1.18 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 183 | 1850000 | 1400582 | --- | 0.48 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 184 | 225000 | 293158 | --- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 185 | 102000 | 30222 | --- | 0.75 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 186 | 1600000 | 567553 | --- | 0.60 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 187 | 129000 | 42646 | --- | 0.28 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 188 | 1850000 | 2079313 | --- | 0.20 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 189 | 80000 | 20280 | -0- | 1.50 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 190 | 12500 | 4900 | -0- | Nil | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 191 | 100000 | 30000 | -0- | 2.5 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 192 | 75000 | 24000 | -0- | 2.5 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
|  |  |  | Total Plant System |  | Field Gas Sweeting |  |  |  |  |  |  | Sulfur Recovery Plant |  |  |  |  | Tail Gas Treating |  |  |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After | Sweetening Process | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 193 | 36000 | 13000 | 0.10 | 0.04 | No | --- | --- | --- | --- | MEA | F60 | --- | --- | --- | --- | --- | --- | --- | --- |
| 194 | --- | --- | --- | --- | Yes | NSul | 1800 | 8.0 | -0- | NSul | F129 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Sweetening Plant. Capacity = 2000 MCFD. Inlet MOL % $CO_2$ = 0.5) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 195 | --- | --- | --- | --- | --- | --- | --- | --- | --- | Amine | F30 | --- | --- | --- | --- | --- | --- | --- | --- |
| 195A |  | 350 | 6.7 | 0.27 | Yes | Amine | 350 | 6.7 | -0- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 195B |  | 2550 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| (Sweetening plant. Capacity = 3000 MCFD) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 196 | 23000 | 17200 | 0.001 | 0.3 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 196A | --- | --- | --- | --- | Yes | IrSp | 260 | 0.08 | 0.0004 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 196B | --- | --- | --- | --- | Yes | IrSp | 1700 | 0.07 | 0.0004 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 196C | --- | --- | --- | --- | Yes | IrSp | 600 | 0.08 | 0.0004 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 196D | --- | --- | --- | --- | Yes | IrSp | 1000 | 0.0004 | 0.0004 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 196E | --- | --- | --- | --- | Yes | IrSp | 2500 | 0.0008 | 0.0004 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 196F | --- | --- | --- | --- | Yes | Amine | 500 | 0.15 | -0- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 197 | 180000 | 118000 | 0.0001 | 0.090 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 197A |  |  |  |  | Yes | IrSp | 6000 | 0.016 | 0.0004 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 197B |  |  |  |  | Yes | Amine | 10000 | 0.020 | 0.0004 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 197C |  |  |  |  | Yes | Sulf | 7000 | 0.0032 | 0.0004 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 198 | 20000 | 5000 | 0.158 | 4.0 | No | --- | --- | --- | --- | DGA | F126 | --- | --- | --- | --- | --- | --- | --- | --- |
| 199 | 1500 | 1500 | Nil | Nil | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 200 | 350000 | 271807 | --- | 0.38 | Yes | (14 Plants) |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 200A | 10000 | 10000 | --- | --- | Yes | Sulf | 10000 | 0.825 | 0.0004 | Sulf | F50 | --- | --- | --- | --- | --- | --- | --- | --- |
| 200B | 4600 | 4600 | --- | --- | Yes | Sulf | 4600 | 0.825 | 0.0004 | Sulf | F100 | --- | --- | --- | --- | --- | --- | --- | --- |
| 200C | 10000 | 10000 | --- | --- | Yes | Sulf | 10000 | 0.825 | 0.0004 | Sulf | F125 | --- | --- | --- | --- | --- | --- | --- | --- |
| 200D | 4600 | 4600 | --- | --- | Yes | Sulf | 4600 | 0.825 | 0.0004 | Sulf | F100 | --- | --- | --- | --- | --- | --- | --- | --- |
| 200E | 10000 | 10000 | --- | --- | Yes | Sulf | 10000 | 0.825 | 0.0004 | Sulf | F125 | --- | --- | --- | --- | --- | --- | --- | --- |
| 200F | 4600 | 4600 | --- | --- | Yes | Sulf | 4600 | 0.825 | 0.0004 | Sulf | F120 | --- | --- | --- | --- | --- | --- | --- | --- |
| 200G | 4600 | 4600 | --- | --- | Yes | Sulf | 4600 | 0.825 | 0.0004 | Sulf | F115 | --- | --- | --- | --- | --- | --- | --- | --- |
| 200H | 35000 | 30000 | --- | --- | Yes | Sulf | 30000 | 1.79 | 0.0004 | Sulf | SR | Claus | 23 | 24.4 | 23.1 | 2910 | Incin | --- | Sales |
| 200J | 25000 | 25000 | --- | --- | Yes | Sulf | 25000 | 0.825 | 0.0004 | Sulf | F100 | --- | --- | --- | --- | --- | --- | --- | --- |
| 200K | 4600 | 4600 | --- | --- | Yes | Sulf | 4600 | 0.825 | 0.0004 | Sulf | F80 | --- | --- | --- | --- | --- | --- | --- | --- |
| 200M | 4600 | 4600 | --- | --- | Yes | Sulf | 4600 | 0.825 | 0.0004 | Sulf | F80 | --- | --- | --- | --- | --- | --- | --- | --- |
| 200N | 4600 | 4600 | --- | --- | Yes | Sulf | 4600 | 0.825 | 0.0004 | Sulf | F70 | --- | --- | --- | --- | --- | --- | --- | --- |
| 200P | 4600 | 4600 | --- | --- | Yes | Sulf | 4600 | 0.825 | 0.0004 | Sulf | F80 | --- | --- | --- | --- | --- | --- | --- | --- |
| 200Q | 10000 | 10000 | --- | --- | Yes | Sulf | 10000 | 0.825 | 0.0004 | Sulf | F150 | --- | --- | --- | --- | --- | --- | --- | --- |

# API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | Total Plant System | | Field Gas Sweeting | | | | | Sweet-ening Proc-ess | Acid Gas Disp | Sulfur Recovery Plant | | | | | | | |
| | | | | | | | | MOL % $H_2S$ | | | | | | | | | Tail Gas Treating | | |
| | | | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Proc-ess Used | Vol (MCFD) | Before | After | | | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 201 | --- | --- | --- | --- | Yes | DEAC | 12000 | Nil | Nil | DEA | V75 | --- | --- | --- | --- | --- | --- | --- | --- |
| ($CO_2$ treating plant. Capacity = 20000 MCFD. Inlet MOL % $CO_2$ = 5.0) | | | | | | | | | | | | | | | | | | | |
| 202 | --- | --- | --- | --- | Yes | Amine | 2000 | 0.04 | Nil | Amine | V75 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 4500 MCFD. Inlet MOL % $CO_2$ = 20.0) | | | | | | | | | | | | | | | | | | | |
| 203 | 10000 | 1000 | Nil | 2.86 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 204 | 15000 | 5606 | 7.6 | 1.6 | No | --- | --- | --- | --- | Sulf | SR | Claus | 35 | 13.45 | 13 | 1010 | Incin | --- | Sales |
| 205 | 100000 | 98114 | 36.3 | 7.0 | No | --- | --- | --- | --- | Sulf | SR | Claus | 1300 | 1120 | 1089 | 69440 | Incin | --- | Sales |
| 206 | 100000 | 35063 | 0.0004 | 1.25 | Yes | --- | 35000 | 0.006 | 0.0004 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| | | | | | | (DEA & MDEA with iron sponge back-up) | | | | | | | | | | | | | |
| 207 | 4000 | 1000 | 0.003 | 1.1 | No | --- | --- | --- | --- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 208 | 31000 | 24000 | -0- | -0- | (UNK-CONTROLLED BY OTHER) | | | | | Ab | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 209 | 1700 | 1854 | --- | 0.72 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 210 | 1200 | 1204 | --- | 0.72 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 211 | 12000 | 6187 | 0.35 | 1.33 | No | --- | --- | --- | --- | Sulf | F200 | --- | --- | --- | --- | --- | --- | --- | --- |
| 212 | 4000 | 1400 | 2.0 | 2.75 | No | --- | --- | --- | --- | F180 | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 213 | 3000 | 1028 | 0.0004 | 1.20 | Yes | IrSp | 1028 | 0.0016 | 0.0004 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 214 | 12000 | 2700 | -0- | 1.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 215 | 15000 | 900 | -0- | 2.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | | | | | | | | Sulfur Recovery Plant | | | | | | | |
| | | | Total Plant System | | Field Gas Sweeting | | | | | | | | | | | | Tail Gas Treating | | |
| | | | | | | | | MOL % $H_2S$ | | | | | | (A) | (B) | (A)-(B) | | | |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Proc-ess Used | Vol (MCFD) | Before | After | Sweet-ening Proc-ess | Acid Gas Disp | Type | Design Cap LT/Day | Avail. Sulf LT/Day | Corres Meas. Rec. LT/Day | Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 216 | 6000 | 2250 | -0- | 0.88 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 217 | 7500 | 3000 | 0.16 | 6.9 | No | --- | --- | --- | --- | DEA | I102 | --- | --- | --- | --- | --- | --- | --- | --- |
| 218 | 30000 | 18000 | 0.47 | 0.3 | No | ADEA | F150 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 219 | 30000 | 18000 | 0.00005- | 0.08 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 220 | 45000 | 30000 | -0- | 0.45 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 221 | --- | --- | --- | --- | Yes | MEAG | Idle | --- | --- | MEAG | SR | 2StCl | 63.25 | 34.78 | 27.04 | 17338 | --- | --- | --- |
| (Treating plant. Capacity = 164,000 MCFD. Process acid gas for others. See Plant Nos. 230, 477 and 479.) | | | | | | | | | | | | | | | | | | | |
| 222 | 120000 | Idle | --- | --- | --- | --- | --- | --- | --- | MEAG | F100 | --- | --- | --- | --- | --- | --- | --- | --- |
| 223 | 100000 | 67430 | 0.24 | 1.22 | No | --- | --- | --- | --- | MEA | F171 | --- | --- | --- | --- | --- | --- | --- | --- |
| 224 | 180000 | 190270 | 0.23 | 1.28 | No | --- | --- | --- | --- | DEA | SR | 3StCl | 22.6 | 16.47 | 9.62 | 15344 | --- | --- | Sales |
| 225 | --- | --- | --- | --- | Yes | MEAG | 9590 | 0.03 | --- | MEAG | F100 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 147,000 MCFD. Inlet MOL % $CO_2$ = 0.16) | | | | | | | | | | | | | | | | | | | |
| 226 | --- | --- | --- | --- | Yes | MEAG | 103690 | 0.22 | --- | MEAG | F100 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 210,000 MCFD. Inlet MOL % $CO_2$ = 0.98) | | | | | | | | | | | | | | | | | | | |
| 227 | --- | --- | --- | --- | Yes | DEA | 29975 | 2.23 | --- | DEA | SR | 3StCl | 24.78 | 25.16 | 23.9 | 2822 | --- | --- | Sales |
| (Treating plant. Capacity = 50,000 MCFD. Inlet MOL % $CO_2$ = 7.57) | | | | | | | | | | | | | | | | | | | |
| 228 | 185000 | 100000 | 0.5 | 0.5 | No | --- | --- | --- | --- | DEA | SR | 3StCl | 38.6 | 20 | 18 | 4480 | Incin | 4480 | TrLiq |
| 229 | 43000 | 34982 | 2.7 | 1.4 | No | --- | --- | --- | --- | MEA | SR | 3StCl | 55 | 32 | 26 | 13440 | --- | --- | Sold |
| | | | | | | | | | | | (Also I100 & F75) | | | | | | | | |
| 230 | 80000 | 88000 | 1.096 | 3.700 | No | --- | --- | --- | --- | --- | SR | (See Plant No. 221.) | | | | --- | --- | --- | --- |
| | | | | | | | | | | | (Also F82) | | | | | | | | |
| 231 | 35000 | 28000 | 1.375 | 1.92 | No | --- | --- | --- | --- | MEA | SR | 3StCl | 26.4 | 22.8 | 18 | 10752 | --- | --- | Sold |
| | | | | | | | | | | | (Also I100 & F75) | | | | | | | | |
| 232 | 60000 | 26000 | -0- | 0.53 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 233 | 19000 | 5000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 234 | 30000 | 28000 | -0- | 0.30 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 235 | 22500 | 9400 | 0.52 | 1.66 | No | --- | --- | --- | --- | DGA | F141 | --- | --- | --- | --- | --- | --- | --- | --- |
| 236 | --- | --- | --- | --- | Yes | Amine | 16000 | 2.0 | Nil | Amine | F | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 20,000 MCFD. Inlet MOL % $CO_2$ = 1.5) | | | | | | | | | | | | | | | | | | | |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | Total Plant System | | Field Gas Sweeting | | | | | Sweetening Process | Acid Gas Disp | Sulfur Recovery Plant | | | | | | | |
| | | | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % $H_2S$ | | | | | | | | | Tail Gas Treating | | |
| | | | | | | | | Before | After | | | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 237 | 45000 | 30500 | 1.1 | 5.05 | No | --- | --- | --- | --- | MEA | SR | 3StCl | 16 | 12.54 | 11.46a | 2419a | --- | --- | Sold |
| (Note a: Col 17 & 18 volumes are operating averages--annual averages are 9.32 and 7213.) | | | | | | | | | | | | | | | | | | | |
| 238 | 4000 | 3600 | 1.21 | 1.9 | No | --- | --- | --- | --- | MEA | I<br>(Tip incinerator) | --- | --- | --- | --- | --- | --- | --- | --- |
| 239 | 7500 | 5050 | -0- | 0.14 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 240 | 30000 | 8000 | 0.30 | 1.45 | No | --- | --- | --- | --- | MEA | F65 | --- | --- | --- | --- | --- | --- | --- | --- |
| 240A | 50000 | 28000 | --- | 0.55 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 241 | 97000 | 130000 | 0.0005 | 1.30 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 242 | 88000 | 60000 | 0.0005 | 1.60 | No | --- | --- | --- | --- | DEAC | V | --- | --- | --- | --- | --- | --- | --- | --- |
| 243 | 85000 | 65000 | -0- | 0.60 | No | --- | --- | --- | --- | DEA | V | --- | --- | --- | --- | --- | --- | --- | --- |
| 244 | 5000 | 750 | -0- | 2.00 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 245 | 125000 | 19000 | 0.20 | 2.07 | No | --- | --- | --- | --- | Amine | F150 | --- | --- | --- | --- | --- | --- | --- | --- |
| 246 | 100000 | 61000 | 0.09 | 0.63 | No | --- | --- | --- | --- | Amine | F135 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Note: Gas is compressed & treated in No. 246 then processed in 245.) | | | | | | | | | | | | | | | | | | | |
| 247 | 15000 | 9000 | Nil | 0.34 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 248 | 94000 | 65110 | 0.16 | 0.9 | No | --- | --- | --- | --- | Amine | F100 | --- | --- | --- | --- | --- | --- | --- | --- |
| 249 | 77000 | 54000 | 0.83 | 2.30 | No | --- | --- | --- | --- | DEA | SR<br>(Also I100) | Claus | 30 | 16.8 | 16.1 | 1568 | --- | --- | Truck |
| 250 | 26000 | 20732 | 0.24 | 0.51 | No | --- | --- | --- | --- | Amine | F75 | --- | --- | --- | --- | --- | --- | --- | --- |
| 251 | 26000 | 26027 | -0- | 0.58 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 252 | 10000 | 4277 | 0.27 | 1.92 | No | --- | --- | --- | --- | Amine | F175 | --- | --- | --- | --- | --- | --- | --- | --- |
| 253 | 42000 | 38400 | 2.5 | 1.0 | No | --- | --- | --- | --- | MEA | SR | 3StCl | 36.0 | 35.5 | 34.1 | 3140 | --- | 3141 | Sales |
| 254 | 220000 | 140000 | 0.3 | 1.6 | No | --- | --- | --- | --- | MEA | SR | Claus | 20 | 11 | 10 | 2240 | --- | --- | --- |
| 255 | 61000 | 54000 | -0- | 0.42 | No | --- | --- | --- | --- | DEA | V20 | --- | --- | --- | --- | --- | --- | --- | --- |

API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
|  |  |  | Total Plant System |  | Field Gas Sweeting |  |  |  |  |  |  | Sulfur Recovery Plant |  |  |  |  |  |  |  |
|  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  | Tail Gas Treating |  |  |
|  |  |  |  |  |  |  |  | MOL % $H_2S$ |  |  |  |  |  |  |  |  |  |  |  |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Proc-ess Used | Vol (MCFD) | Before | After | Sweet-ening Proc-ess | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 256 | 10000 | 4000 | 0.003 | 1.1 | No | --- | --- | --- | --- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 257 | 27000 | 3000 | 0.012 | 1.4 | Yes | IrSp | 600 | 0.012 | 0-Fuel | MDEA | F212 | --- | --- | --- | --- | --- | --- | --- | --- |
| 258 | 110000 | 53700 | 3.85 | 1.50 | No | --- | --- | --- | --- | MEA | SR | Claus | 157 | 87 | 82.9 | 9020 | None | --- | Sales |
|  |  |  |  |  |  |  |  |  |  |  | (Also I75) |  |  |  |  |  |  |  |  |
| 259 | 8000 | 6000 | 1.65 | 1.5 | No | --- | --- | --- | --- | MEA | I200 | --- | --- | --- | --- | --- | --- | --- | --- |
|  |  |  |  |  |  |  |  |  |  |  | (Tip incinerator) |  |  |  |  |  |  |  |  |
| 260 | 24000 | 16780 | 9.50 | 2.00 | No | --- | --- | --- | --- | DEA | SR | ClCBA | 72 | 60.0 | 58.8 | 2688 | CBA | --- | Truck |
|  |  |  |  |  |  |  |  |  |  |  | (Also I195) |  |  |  |  |  |  |  |  |
| 261 | 2500 | 900 | 0.0022 | 1.05 | No | --- | --- | --- | --- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | Bury |
| 262 | 6000 | 1500 | -0- | 0.8 | Yes | Amine | 1500 | 0.5 | -0- | Amine | F200 | --- | --- | --- | --- | --- | --- | --- | --- |
|  |  |  |  |  |  | (At plant inlet) |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 263 | 12000 | --- | 0.0004- | 0.1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 264 | 5000 | 4000 | 0.0004- | 0.1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 265 | 2000 | 1000 | 0.0004- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | | | | | | | | Sulfur Recovery Plant | | | | | | | |
| | | | Total Plant System | | Field Gas Sweeting | | | MOL % $H_2S$ | | | | | | | | | Tail Gas Treating | | |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | Before | After | Sweetening Process | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 266 | 10000 | 4000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 267 | 1000 | 600 | -0- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 268 | 60000 | 35000 | -0- | 0.15 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 269 | 45000 | 45000 | -0- | 0.31 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 270 | 75000 | 57000 | -0- | 0.16 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 271 | 200000 | 160000 | -0- | 1.0 | No | --- | --- | --- | --- | DEA | V | --- | --- | --- | --- | --- | --- | --- | --- |
| 272 | 26000 | 8500 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 273 | 20000 | 3400 | -0- | 0.23 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 274 | 6000 | 6400 | -0- | 0.62 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 275 | 66000 | 36000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 276 | 250000 | 180000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 277 | 32000 | 23000 | -0- | 0.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 278 | 20000 | 10695 | -0- | 0.47 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 279 | 15000 | 8000 | -0- | 0.47 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 280 | 12000 | 6000 | 0.0004- | 0.37 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 281 | 15000 | 11000 | 0.0004- | 0.47 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 282 | 15000 | 6000 | 0.001 | 0.46 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 283 | 20000 | 10000 | 0.0004- | 0.19 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 284 | 3000 | 300 | Nil | 0.20 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 285 | 12000 | 1700 | Nil | 0.45 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 286 | 17000 | 10400 | Nil | 0.25 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 287 | 3000 | 1800 | Nil | 0.40 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 288 | 17500 | 10800 | Nil | 0.20 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 289 | 45000 | 45000 | Nil | 0.35 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 290 | 200000 | 120000 | Nil | 0.40 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 291 | 3000 | 2000 | --- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 292 | 3000 | 700 | --- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 293 | 30000 | --- | --- | --- | --- | --- | --- | --- | --- | MEA | F60 | --- | --- | --- | --- | --- | --- | --- | --- |
| 293A | | 28000 | Trace | 0.09 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 293B | | 2000 | 0.44 | 0.19 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 294 | 75000 | 55000 | Trace | 0.09 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 295 | 15000 | 5000 | Trace | 1.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | Total Plant System | | Field Gas Sweeting | | | | | Sweetening Process | Acid Gas Disp | Sulfur Recovery Plant | | | | | | | |
| | | | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % $H_2S$ | | | | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Tail Gas Treating | | |
| | | | | | | | | Before | After | | | | | | | | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 296 | 50000 | 19000 | -0- | 6.86 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 297 | 150000 | 210000 | 0.0004 | 0.49 | No | --- | --- | --- | --- | MEA | F80 | --- | --- | --- | --- | --- | --- | --- | --- |
| 298 | 56000 | 55000 | 0.0006 | 0.4 | No | --- | --- | --- | --- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 299 | 15000 | Startup | 0.0004 | 1.22 | No | --- | --- | --- | --- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 300 | 25000 | 15000 | 0.0004 | 0.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 301 | 150000 | 35000 | 0.0004 | 0.5 | No | --- | --- | --- | --- | MEA | F90 | --- | --- | --- | --- | --- | --- | --- | --- |
| | | (Startup conditions. Estimate 82000 MCFD thruput by late 1982 with similar feed.) | | | | | | | | | | | | | | | | | |
| 302 | 10000 | 4500 | 0.0002- | 0.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 303 | 27000 | 11000 | 0.0002- | 0.5 | No | --- | --- | --- | --- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 304 | 100000 | | | | | | | | | | | | | | | | | | |
| 304A | | 2500 | 0.42 | 0.5 | No | --- | --- | --- | --- | MEA | F50 | --- | --- | --- | --- | --- | --- | --- | --- |
| 304B | | 19500 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 305 | 75000 | 45000 | --- | 0.19 | No. | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 306 | 10000 | 5000 | 0.0003 | 0.43 | No | --- | --- | --- | --- | IrSp | SR | WCFB | 0.008 | 0.0006 | 0.0006 | -0- | --- | --- | LaF1 |
| 307 | 60000 | 49700 | -0- | 0.62 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 308 | 24000 | 15900 | -0- | 0.65 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 309 | 18500 | 3500 | 0.0025 | 0.53 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 310 | 50000 | 42500 | -0- | 0.81 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 311 | 20000 | 4300 | -0- | 0.36 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 312 | 70000 | 37400 | -0- | 0.84 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 313 | 11000 | 1400 | -0- | 0.82 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 314 | 45000 | 32000 | -0- | 0.10 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 315 | 70000 | 35000 | 1.58 | 0.21 | No | --- | --- | --- | --- | MEAA | F60 | --- | --- | --- | --- | --- | --- | --- | --- |
| | | | (2500 MCFD sour gas) | | | | | | | | | | | | | | | | |
| 316 | 14000 | 12000 | -0- | 0.09 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 317 | 10000 | 8000 | -0- | 0.09 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 318 | 3000 | 25000 | -0- | 0.09 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 319 | 65000 | 29000 | 0.004 | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 320 | 20000 | 1250 | -0- | -0- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
|  |  |  | Total Plant System |  | Field Gas Sweeting |  |  |  |  |  |  | Sulfur Recovery Plant |  |  |  |  | Tail Gas Treating |  |  |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After | Sweetening Process | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 321 | 15000 | 9000 | --- | 0.73 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| √322 | 80000 | 69000 | 0.001 | 0.47 | No | --- | --- | --- | --- | DEA | 160 | --- | --- | --- | --- | --- | --- | --- | --- |
| 323 | 7500 | 3000 | Nil | 0.10 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 324 | 45000 | 15500 | None | 2.20 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 325 | 23000 | 12000 | Nil | 0.17 | No | (BY OTHERS) |  | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| √326 | --- | --- | --- | --- | Yes | IrSp | 24000 | 0.3 | Nil | IrSp | None | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. | Capacity = 30,000 MCFD. |  |  | Inlet MOL % $CO_2$ = 0) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 327 | 290000 | 233000 | -0- | 0.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 328 | 40000 | 19000 | -0- | 0.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 329 | 90000 | 43500 | -0- | 0.32 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 330 | 30000 | 15000 | --- | 0.40 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 330A | 25000 | 23000 | --- | 0.40 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 331 | 75000 | --- | --- | 0.7 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 332 | 30000 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 332A |  | 28000 | --- | 0.20 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 332B |  | 5000 | --- | 0.20 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 333 | 36000 | 30000 | -0- | 0.25 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 334 | 10000 | 8000 | -0- | 1.25 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 335 | 30000 | 30000 | Nil | 0.21 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 336 | 100000 | --- | --- | --- | --- | --- | --- | --- | --- | Amine | F170 | --- | --- | --- | --- | --- | --- | --- | --- |
| 336A |  | 57000 | 0.046 | 0.588 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 336B |  | 2000 | Nil | Nil | Yes | Amine | 2000 | 0.42 | Nil | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 336C |  | 2000 | 0.01 | 0.39 | Yes | Amine | 2000 | 0.51 | 0.01 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 336D |  | 3000 | Nil | 0.41 | Yes | Amine | 3000 | 0.80 | Nil | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 336E |  | 2500 | Nil | 0.07 | Yes | Amine | 2500 | 0.47 | Nil | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 336F |  | 2500 | Nil | 0.28 | Yes | Amine | 2500 | 0.38 | Nil | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 336G |  | 1000 | Nil | Nil | Yes | Amine | 1000 | 0.68 | Nil | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | | | | | | | | Sulfur Recovery Plant | | | | | | | |
| | | | Total Plant System | | Field Gas Sweeting | | | | | | | | | | | | Tail Gas Treating | | |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Proc-ess Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After | Sweet-ening Proc-ess | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 337 | 15000 | 10000 | -0- | 0.35 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 338 | 50000 | 16751 | -0- | 0.71 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 339 | 40000 | 40000 | -0- | 0.79 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 340 | 59540 | --- | --- | --- | No | | | | | | | | | | | | | | |
| 340A | | 20000 | 0.006 | 0.35 | --- | --- | --- | --- | --- | Amine | F119 | --- | --- | --- | --- | --- | --- | --- | --- |
| 340B | | 35627 | -0- | 0.35 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 341 | 175000 | 41600 | -0- | 0.33 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 342 | 80000 | 26500 | --- | 0.89 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 343 | 55000 | 31000 | --- | 0.45 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 344 | 150000 | 74800 | --- | 0.64 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 345 | 25000 | 12200 | --- | 0.45 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 346 | 39000 | 21400 | --- | 0.39 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 347 | 20000 | 4300 | --- | 0.34 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 348 | 20000 | 15000 | -0- | 0.40 | No | --- | --- | --- | --- | DEA | V30 | (DEA removes $CO_2$ from liquid product.) | | | | | | --- | --- |
| 349 | 75000 | 51730 | Nil | 0.67 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 350 | 9000 | 6800 | -0- | 0.15 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Plant or Stream No. | Current Capacity (MCFD) | Current Throughput (MCFD) | Total Plant System |  | Field Gas Sweeting |  |  |  |  | Sweetening Process | Acid Gas Disp | Sulfur Recovery Plant |  |  |  |  | Tail Gas Treating |  |  |
|  |  |  | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After |  |  | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 351 | 7000 | 800 | --- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 352 | 60000 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 352A |  | 1500 | 0.0004 | --- | Yes | Amine | 1500 | 0.18 | 0.0004 |  | Amine |  | F120 | --- | --- | --- | --- | --- | --- |
| 352B |  | 33500 | --- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 353 | 10000 | 7500 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 354 | 87500 | 54000 | Trace | 2.65 | Yes | (CO$_2$ removal only) |  |  | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 355 | 10000 | 3000 | --- | 0.16 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 356 | 16000 | 10000 | -0- | 1.48 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 357 | 8000 | 2000 | -0- | 1.38 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 358 | 600 | 250 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 359 | 5000 | 500 | -0- | 0.06 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 360 | 370000 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 360A |  | 170000 | -0- | 0.9 | No | --- | --- | --- | --- | MEA | V | --- | --- | --- | --- | --- | --- | --- | --- |
| 360B |  | 10000 | 0.25 | 0.4 | No | --- | --- | --- | --- | MEA |  |  |  |  |  |  |  |  |  |
| 361 | 130000 | 125000 | 0.25 | 0.4 | No | --- | --- | --- | --- | MEA | SR | Claus | 30 | 12.7 | 12.1 | 1344 | Incin | 1344 | Sales |

Note for 360 & 361: Refinery acid gas is also charged to Sulf. Plant and some acid gas from Gas Plant amine system is charged to Acid Plant. Numbers in Col. 15-20 assume only acid gas from Gas Plant is charged to Sulf. Plant and that none of this gas is charged to Acid Plant.

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| 362 | 160000 | 100000 | -0- | 0.40 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 363 | 100000 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 363A |  | 50000 | -0- | 0.30 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 363B |  | 20000 | 0.05 | 1.00 | No | --- | --- | --- | --- | DEA | F5 | --- | --- | --- | --- | --- | --- | --- | --- |
| 364 | 75000 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 364A |  | 69000 | 0.46 | 0.61 | No | --- | --- | --- | --- | DEA | SR<br>Also | 3StCl<br>I160 | 25 | 12 | 11.3 | 1600 | --- | --- | Sales |
| 364B |  | 6000 | --- | 0.37 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
|  |  |  | Total Plant System |  | Field Gas Sweeting |  |  |  |  |  |  | Sulfur Recovery Plant |  |  |  |  | Tail Gas Treating |  |  |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After | Sweetening Process | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 365 | 75000 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 365A |  | 22000 | --- | Trace | Yes | DEA | 22000 | 0.30 | --- | --- | I150 | --- | --- | --- | --- | --- | --- | --- | --- |
| 365B |  | 12000 | --- | 0.03 | Yes | DGA | 12000 | 11.00 | --- | --- | SR | 3StCl | UNKNOWN | OP BY | OTHERS |  | --- | --- | Sales |
| 365C |  | 41000 | --- | 0.02 | Yes | DEA | 41000 | 4.42 | --- | --- | SR | 3StCl | UNKNOWN | OP BY | OTHERS |  | --- | --- | Sales |
| 366 | 125000 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 366A |  | 70000 | --- | 1.70 | Yes | MDEA | 70000 | 0.02 | --- | --- | I87 | --- | --- | --- | --- | --- | --- | --- | --- |
| 366B |  | 55000 | --- | 1.29 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 367 | 26000 | 17100 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 368 | 4000 | 1600 | -0- | 0.61 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 369 | 30000 | 4900 | -0- | 0.39 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 370 | 18000 | 14700 | -0- | 0.49 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 371 | 14000 | 12400 | -0- | 0.15 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 372 | 50000 | 40500 | -0- | 0.40 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 373 | 20000 | 16200 | -0- | 2.63 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 374 | 96000 | 12400 | -0- | 0.55 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 375 | 20000 | 18000 | -0- | 0.43 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 376 | 7000 | 4700 | -0- | 0.46 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 377 | 10000 | 3000 | -0- | 0.69 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 378 | 5000 | 5500 | -0- | 1.04 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 379 | 80000 | 74000 | -0- | 0.47 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 380 | 65000 | 68000 | -0- | 1.54 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 381 | 30000 | 17800 | -0- | 0.45 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 382 | 15000 | 7200 | -0- | 0.50 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 383 | 7500 | 4900 | -0- | 0.40 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 384 | --- | --- | --- | --- | Yes | MEA | 40000 | 0.35 | --- | MEA | I150 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Amine plant. Capacity = 60,250 MCFD. Inlet MOL % $CO_2$ = 0.82) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 385 | --- | --- | --- | --- | Yes | DEA | 107360 | 1.03 | --- | DEA | SR | 3StCl | 142.5 | 41.62 | 31.13 | 23497 | --- | --- | Sales |
| (Treating plant. Capacity = 155,000 MCFD. Inlet MOL % $CO_2$ = 0.60) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | Total Plant System |  | Field Gas Sweeting |  |  |  |  | Sweetening Process | Acid Gas Disp | Sulfur Recovery Plant |  |  |  |  |  |  |  |
|  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  | Tail Gas Treating |  |  |
|  |  |  | H₂S (MOL%) | CO₂ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % H₂S Before | MOL % H₂S After |  |  | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 386 | --- | --- | --- | --- | Yes | DEA | 15230 | 6.29 | --- | DEA | SR | 3StCl | 86.6 | 36.06 | 35.59 | 1053 | --- | --- | Sales |
| (Treating plant. Capacity = 54,000 MCFD. Inlet MOL % $CO_2$= 2.44) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 387 | --- | --- | --- | --- | Yes | MEAG | 12140 | 0.67 | --- | MEAG | F80 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 40,000 MCFD. Inlet MOL % $CO_2$ = 1.35) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 388 | --- | --- | --- | --- | Yes | DGA | 282490 | --- | --- | DGA | F100 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 320,000 MCFD. Inlet MOL % $CO_2$ = 1.61) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 389 | --- | --- | --- | --- | Yes | PrC | 29820 | -0- | -0- | PrC | V80 | --- | --- | --- | --- | --- | --- | --- | --- |
| ($CO_2$ treating plant. Capacity = 80,000 MCFD. Inlet MOL % $CO_2$ = 27.4) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 390 | 30000 | 15840 | 0.1 | 0.52 | No | --- | --- | --- | --- | MEAG | F100 | --- | --- | --- | --- | --- | --- | --- | --- |
| 391 | --- | --- | --- | --- | Yes | PrC | 124860 | -0- | -0- | PrC | V100 | --- | --- | --- | --- | --- | --- | --- | --- |
| ($CO_2$ treating plant. Capacity = 240,000 MCFD. Inlet MOL % $CO_2$ = 37.48) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 392 | --- | --- | --- | --- | Yes | DGA | 184680 | 0.06 | --- | DGA | I100 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 285,000 MCFD. Inlet MOL % $CO_2$ = 2.40) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 393 | --- | --- | --- | --- | Yes | DGA | 29370 | 0.41 | --- | DGA | I120 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 150,000 MCFD. Inlet MOL % $CO_2$ = 10.79) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 394 | --- | --- | --- | --- | Yes | MEAG | 21910 | 0.05 | --- | MEAG | F90 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 30,000 MCFD. Inlet MOL % $CO_2$ = 1.48) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 395 | 15000 | 11000 | -0- | -0- |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 396 | 25000 | 8000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 397 | 150000 | 60000 | -0- | 1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 398 | 18000 |  |  |  | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 398A |  | 9000 | 0.15 | 29 | No | --- | --- | --- | --- | MEA | I100 (Incin when tert. recovery recomp. is down) |  |  |  |  |  |  |  |  |
| 398B |  | 6500 | 0.06 | 1.3 | No | --- | --- | --- | --- | MEA | F0 | --- | --- | --- | --- | --- | --- | --- | --- |
| 399 | 5000 | 2000 | -0- | 0.6 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | Total Plant System | | Field Gas Sweeting | | | | | | | Sulfur Recovery Plant | | | | | Tail Gas Treating | | |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After | Sweetening Process | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 400 | 5000 | 2200 | Nil | --- | Yes | IrSp | 2200 | Nil | -0- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 401 | 5000 | 2400 | -0- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 402 | 20000 | 20000 | 5 | 6 | No | --- | --- | --- | --- | DGA | SR | 2StCl | 33.4 | 27 | 24 | 6720 | Incin | 6720 | TrLiq |
| 403 | 3000 | 3000 | 0.0013 | 0.92 | No | --- | --- | --- | --- | MEA | F60 | --- | --- | --- | --- | --- | --- | --- | --- |
| 404 | 120000 | | | | | | | | | | | | | | | | | | |
| 4040 | (Offshore) | 90000 | -0- | 0.8 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 404L | (Onshore) | 20000 | -0- | 0.8 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 405 | 60000 | 15000 | -0- | 0.51 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 406 | 110000 | 37000 | -0- | 1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 407 | 3500 | 3500 | 0.0006 | 2 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 408 | 50000 | 15000 | -0- | 1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 409 | 1000 | 1000 | -0- | 2 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 410 | 10000 | 10000 | 0.0034 | 1 | No | --- | --- | --- | --- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| | | | (Tail gate iron sponge unit on sales gas after plant processing.) | | | | | | | | | | | | | | | | |
| 411 | 3500 | 2500 | -0- | 0.6 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 412 | 32000 | | | | | | | | | | | | | | | | | | |
| 412A | | 630 | 0.294 | 0.146 | No | --- | --- | --- | --- | Amine | F70 | --- | --- | --- | --- | --- | --- | --- | --- |
| 412B | | 26903 | 0.065 | 0.146 | No | --- | --- | --- | --- | AmC | V25 | --- | --- | --- | --- | --- | --- | --- | --- |
| 413 | 15000 | 8000 | 0.040 | 0.520 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 414 | 24000 | 24834 | 0.040 | 0.055 | No | --- | --- | --- | --- | AmC | V25 | --- | --- | --- | --- | --- | --- | --- | --- |
| 415 | 8000 | 3070 | -0- | 2.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 416 | 15000 | 7000 | 0.001 | 0.55 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 417 | 50000 | 22000 | 0.0004- | 0.39 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 418 | 56000 | 16000 | 0.0004- | 0.34 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 419 | 18000 | 2000 | 0.0004- | 0.80 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 420 | 35000 | 10000 | 0.015 | 0.73 | No | --- | --- | --- | --- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 421 | 12000 | 8000 | 0.001 | 0.50 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 422 | 90000 | 61000 | 0.0004- | 2.86 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 423 | 15000 | 2500 | 0.0004- | 1.18 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 424 | 6500 | 3000 | 0.0004- | 0.61 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 425 | 50000 | 12000 | 0.0004- | 1.27 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 426 | 40000 | 27000 | 0.0004- | 0.58 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Plant or Stream No. | Current Capacity (MCFD) | Current Throughput (MCFD) | Total Plant System | | Field Gas Sweeting | | | | | Sweetening Process | Acid Gas Disp | Sulfur Recovery Plant | | | | | | | |
| | | | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % $H_2S$ | | | | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Tail Gas Treating | | |
| | | | | | | | | Before | After | | | | | | | | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 427 | 180000 | | | | | | | | | | | | | | | | | | |
| 427A | | 55000a | 0.07 | 3.7 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 427B | | 65000a | 0.06 | 4.8 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 427C | | 60000a | 0.06 | 5.7 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 427D | | 50000b | 0.05 | 6.8 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 427E | | 25000c | 0.08 | 3.7 | --- | --- | --- | --- | --- | MEA | F100 | --- | --- | --- | --- | --- | --- | --- | --- |
| | | Note: a. Gas Recycled to reservoir.<br>b. Gas bypasses process and is recycled to the reservoir.<br>c. Fuel gas is residue from stream A processing that is treated through an amine system. | | | | | | | | | | | | | | | | | |
| 428 | 8000 | 6000 | 1.7 | 6.9 | Yes | MEA | 6000 | 1.7 | 0.0004 | MEA | SR | 2StCl | 10 | 3.8 | 3.5 | 672 | --- | --- | --- |
| 429 | --- | --- | --- | --- | Yes | MDEA | 30000 | 0.0037 | 0.0004 | MDEA | I60 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Sweetening plant. Capacity = 200,000 MCFD. Inlet MOL % $CO_2$ = 9.74. Startup--expect throughput at capacity by mid 1983.) | | | | | | | | | | | | | | | | | | | |
| 430 | 20000 | | | | | | | | | | | | | | | | | | |
| 430A | | 6000 | 0.0025 | 0.20 | No | --- | --- | --- | --- | DEA | I50 | --- | --- | --- | --- | --- | --- | --- | --- |
| 430B | | 1500 | Nil | 0.20 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 430C | | 700 | Nil | 0.40 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 431 | 50000 | 53000 | 0.0035 | 0.50 | No | --- | --- | --- | --- | DEA | V60 | --- | --- | --- | --- | --- | --- | --- | --- |
| 432 | --- | --- | --- | --- | Yes | KCO | 103000 | 0.045 | --- | KCO | I68<br>(Also V125) | --- | --- | --- | --- | --- | --- | --- | --- |
| ($CO_2$ treating plant. Capacity = 105,000 MCFD. Inlet MOL % $CO_2$ = 40.0) | | | | | | | | | | | | | | | | | | | |
| 433 | 90000 | --- | 0.0003 | 0.50 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 433A | | 64000 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 433B | | 6000 | --- | --- | No | SlSw | 6000 | 0.0015 | 0.0003 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 434 | 10000 | 2000 | Nil | 0.30 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 435 | 150000 | 65000 | --- | --- | Yes (See Plant No. 432) | | | | | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 436 | 75000 | 20200 | Nil | 0.50 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 437 | 35000 | 15100 | Nil | 3.20 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 438 | 140000 | 22400 | Nil | 0.40 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 439 | 45000 | 16100 | Nil | 2.60 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 440 | 10000 | 7000 | Nil | 0.90 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | Total Plant System | | Field Gas Sweeting | | | | | Sweetening Process | Acid Gas Disp | Sulfur Recovery Plant | | | | | | | |
| | | | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % $H_2S$ | | | | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Tail Gas Treating | | |
| | | | | | | | | Before | After | | | | | | | | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 441 | 10000 | 4000 | --- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 442 | 3000 | 500 | --- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 443 | --- | --- | --- | --- | Yes | Amine | 600 | 0.5 | --- | Amine | I30 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 2000 MCFD. Inlet MOL % $CO_2$ = 1.5) | | | | | | | | | | | | | | | | | | | |
| 444 | 55000 | 17800 | -0- | 0.31 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 445 | 300000 | 140000 | 0.033 | 0.24 | No | --- | --- | --- | --- | MEA | F150 | --- | --- | --- | --- | --- | --- | --- | --- |
| | | | | | | | | | | (MEA regeneration at inlet and product treating) | | | | | | | | | |
| 446 | 1500 | 1300 | -0- | 0.27 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 447 | 1000 | 750 | -0- | 1.01 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 448 | 60000 | 23000 | -0- | 1.97 | No | --- | --- | --- | --- | (Inlet-None. Product treating extracted to residue pipeline) | | | | | | | | | |
| 449 | 320000 | 139600 | 0.00035 | 0.207 | No | --- | --- | --- | --- | (Inlet-None. Product treating extracted to residue pipeline) | | | | | | | | | |
| 450 | 10000 | 5400 | -0- | 1.32 | No | --- | --- | --- | --- | DEAC | V30 | --- | --- | --- | --- | --- | --- | --- | --- |
| | | | | | | | | | | (DEA regen. to 0.15 $CO_2$ MOL%) | | | | | | | | | |
| 451 | 20000 | 16900 | -0- | 2.92 | No | --- | --- | --- | --- | DEAC | V22 | --- | --- | --- | --- | --- | --- | --- | --- |
| | | | | | | | | | | (DEA regen. to 0.21 $CO_2$ MOL%) | | | | | | | | | |
| 452 | 90000 | | | | | | | | | | | | | | | | | | |
| 452A | | 80000 | 0.0011 | 2.60 | No | --- | --- | --- | --- | DEA | F60 | --- | --- | --- | --- | --- | --- | --- | --- |
| 452B | | 18200 | 1.0 | 2.75 | Yes | DEA | 18200 | 1.0 | 0.001- | --- | F100 | --- | --- | --- | --- | --- | --- | --- | --- |
| 452C | | 700 | 0.69 | 1.75 | Yes | DEA | 700 | 0.07 | 0.001- | --- | F60 | --- | --- | --- | --- | --- | --- | --- | --- |
| 453 | 12000 | Startup | -0- | 1.27 | No | --- | --- | --- | --- | DEAC | V30 | --- | --- | --- | --- | --- | --- | --- | --- |
| | | | | | | | | | | (DEA regen. to 0.20 $CO_2$ MOL%) | | | | | | | | | |
| 454 | 28000 | 14500 | 0.003 | 0.45 | No | --- | --- | --- | --- | DEA | F120 | --- | --- | --- | --- | --- | --- | --- | --- |
| | | | | | | | | | | (Inlet. DEA regen. prod. treating IrSp--disposed not regen.) | | | | | | | | | |
| 455 | 320000 | 121600 | -0- | 0.15 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 456 | 95000 | 105700 | -0- | 1.30 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 457 | 125000 | 102000 | 0.02 | 0.79 | No | --- | --- | --- | --- | DEA | F150 | --- | --- | --- | --- | --- | --- | --- | --- |
| | | | | | | | | | | (Inlet-DEA Regen. Product-DEA Regen. Liq. Contactor) | | | | | | | | | |
| 458 | 25000 | 11000 | -0- | 1.71 | No | --- | --- | --- | --- | DEAC | V30 | --- | --- | --- | --- | --- | --- | --- | --- |
| | | | | | | | | | | (DEA Regen. to 0.27 $CO_2$ MOL %) | | | | | | | | | |
| 459 | 255000 | 65000 | -0- | 2.50 | No | --- | --- | --- | --- | MEA | V50 | --- | --- | --- | --- | --- | --- | --- | --- |
| | | | | | | | | | | (Inlet. None. Outlet. Ethane treat.--MEA regen.) | | | | | | | | | |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | Total Plant System | | Field Gas Sweeting | | | | | | | Sulfur Recovery Plant | | | | | Tail Gas Treating | | |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After | Sweetening Process | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 460 | 65000 | 37000 | Trace | 0.03 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 461 | 120000 | 80000 | Trace | 0.06 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 462 | 35000 | 10000 | Trace | 0.05 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 463 | 40000 | 40000 | Trace | 1.02 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 464 | 150000 | 75000 | Trace | 0.38 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 465 | 35000 | --- | --- | --- | --- | --- | --- | --- | --- | MEA | F60 | --- | --- | --- | --- | --- | --- | --- | --- |
| 465A | | 11000 | Trace | 0.08 | Yes | MEA | 1500 | 1.07 | Trace | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 465B | | 5000 | 0.96 | 0.08 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 466 | 150000 | 130000 | -0- | 0.44 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 467 | 200000 | | | | | | | | | | | | | | | | | | |
| 467A | | 75000 | -0- | 1.20 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 467B | | 160000 | -0- | 1.40 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 468 | 250000 | 225000 | -0- | 1.25 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 469 | 85000 | 65000 | -0- | 0.76 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 470 | 190000 | 80000 | -0- | 0.70 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 471 | 40000 | 40000 | -0- | 0.71 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 472 | 20000 | 16000 | -0- | 0.50 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 473 | 15000 | 15000 | -0- | 0.48 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 474 | 15000 | 9000 | -0- | 0.44 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 475 | 7000 | 3000 | -0- | 0.21 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 476 | 115000 | 23500 | 2.04 | 0.716 | No | --- | --- | --- | --- | MEA | SR<br>(Also I129 & F140) | 1StCl | Unk | 17.83 | 10.47 | 16486 | --- | --- | Sold |
| 477 | 20000 | --- | --- | --- | --- | --- | --- | --- | --- | MEA | SR | (See Plant No. 221) | | | | --- | --- | --- | --- |
| 477A | | 4290 | 0.033 | 0.445 | No | | | | | | | | | | | | | | |
| 477B | | 15070 | 7.36 | 3.024 | No | | | | | | | | | | | | | | |
| 478 | 55000 | 45000 | 4.209 | 5.922 | No | --- | --- | --- | --- | DEA | SR<br>(Also I215 & F82) | 3StCl | 60 | 73 | 63 | 22400 | --- | --- | Sold |
| 479 | --- | --- | --- | --- | --- | --- | --- | --- | --- | MEA | SR | (See Plant No. 221) | | | --- | --- | --- | --- | --- |
| 479A | 80000 | 15150 | 1.19 | 0.80 | No | | | | | | | | | | | | | | |
| 479B | 80000 | 80463 | -0- | 0.04 | No | | | | | | | | | | | | | | |
| 480 | 250000 | 105000 | -0- | 28.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 481 | 15000 | 11000 | -0- | 0.889 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | | | | | | | | Sulfur Recovery Plant | | | | | | | |
| | | | Total Plant System | | Field Gas Sweeting | | | | | | | | | | | | Tail Gas Treating | | |
| | | | | | | | | MOL % $H_2S$ | | | | | | | | | | | |
| Plant or Stream No. | Current Capacity (MCFD) | Current Throughput (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | Before | After | Sweetening Process | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 483 | 55000 | 23000 | -0- | 1.509 | No | --- | --- | --- | --- | DEAC | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| ✓484 | 75000 | 98000 | 0.0007 | 2.5 | No | --- | --- | --- | --- | MDEA | V75 | --- | --- | --- | --- | --- | --- | --- | --- |
| ✓485 | 12000 | 6500 | 1.199 | 3.326 | No | --- | --- | --- | --- | MEA | I150 | --- | --- | --- | --- | --- | --- | --- | --- |
| 486 | 5000 | 1800 | -0- | 2.0 | No | --- | --- | --- | --- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 487 | 10000 | 1000 | -0- | 2.0 | No | --- | --- | --- | --- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| ✓488 | 10000 | 8500 | 5 | 2.3 | No | --- | --- | --- | --- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 489 | 20000 | 7500 | -0- | 1.300 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 490 | 30000 | 26000 | --- | 0.1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 491 | 40000 | 32000 | --- | 0.411 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 492 | 5000 | 2000 | 4.8 | 1.5 | No | --- | --- | --- | --- | --- | SR | 2StCl | --- | 2.15 | 2 | 336 | --- | 250 | Sales |
| 493 | 100000 | 18000 | --- | 0.06 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| ✓494 | 13000 | 9500 | 0.20 | 0.40 | No | --- | --- | --- | --- | MEA | F50 | --- | --- | --- | --- | --- | --- | --- | --- |
| ✓495 | 19000 | 2100 | 0.50 | 2.0 | No | --- | --- | --- | --- | MEA | F50 | --- | --- | --- | --- | --- | --- | --- | --- |
| 496 | 40000 | 10000 | 0.001 | 1.98 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 497 | 15000 | 8000 | 0.001 | 0.80 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 498 | 250000 | 55297 | 0.03 | 6.8 | Yes | Sulf | 55297 | 0.03 | 0.0004 | Sulf | I185 | --- | --- | --- | --- | --- | --- | --- | --- |
| 499 | 70000 | 28493 | 0.48 | 1.39 | Yes | DEA | 28709 | 0.48 | 0.0004 | DEA | SR | 2StCl | 10.7 | 5.142 | -0- | 11518 | --- | --- | Sales |
| (Note: Claus plant down due to vol. lower than the min. turndown vol. Flaring acid gases from DEA plant. Flare stack is 100 ft.) | | | | | | | | | | | | | | | | | | | |
| 500 | 12000 | 2440 | 5.0 | 7.0 | Yes | DEA | 2990 | 5.00 | 0.0004 | DEA | SR | 3StCL | 25.5 | 4.586 | 1.201 | 7582 | --- | --- | SBP |
| (Note: Claus plant down for repair and lateral replacement. Sulfur production figures given are for 91 days in 1981.) | | | | | | | | | | | | | | | | | | | |
| ✓501 | 40000 | 24000 | 0.048 | 0.05 | Yes | MEA | 24000 | 0.048 | 0.0004 | MEA | F49 | --- | --- | --- | --- | --- | --- | --- | --- |
| 502 | 240000 | 50744 | -0- | 0.537 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 503 | 240000 | 107858 | -0- | 0.092 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 504 | 50000 | 43000 | Trace | 4.68 | Yes | IrSp<br>(IrSp not in service.) | 38500 | 0.0006 | 0.0006 | DEA | V50 | --- | --- | --- | --- | --- | --- | --- | --- |
| 505 | 15000 | 9400 | 3.80 | 5.70 | No | --- | --- | --- | --- | DGA | SR<br>(Also I170) | Claus | 30 | 13 | 12.35 | 1430 | Incin | --- | Sales |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | Total Plant System | | Field Gas Sweeting | | | | | | | Sulfur Recovery Plant | | | | | Tail Gas Treating | | |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Proc-ess Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After | Sweet-ening Proc-ess | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| ✓506 | 5000 | 5800 | 0.0016- | 2.42 | No | --- | --- | --- | --- | DEA | I25 | --- | --- | --- | --- | --- | --- | --- | --- |
| ✓507 | --- | --- | --- | --- | Yes | DEA | 4000 | 0.02 | --- | DEA | I30 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 18,000 MCFD. Inlet MOL % $CO_2$ = 8.60. Lean gas. Goes direct to pipeline with no liquid recovery.) | | | | | | | | | | | | | | | | | | | |
| ✓508 | --- | --- | --- | --- | Yes | DEA | 43000 | 0.06 | --- | DEA | I76 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 40,000 MCFD. Inlet MOL % $CO_2$ = 7.50. Lean gas. Goes direct to pipeline with no liquid recovery.) | | | | | | | | | | | | | | | | | | | |
| 509 | 15000 | 15000 | 0.0016- | 0.30 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 510 | 200000 | | | | | | | | | | | | | | | | | | |
| 5100 | (Offshore) | 149000 | 0.0016- | 0.47 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| ✓511 | 50000 | 13000 | 0.004 | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 512 | 95000 | 4000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 513 | 21000 | 13000 | -0- | -0- | Yes | IrSp | 28000 | 0.003 | -0- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 514 | 75000 | 24000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 515 | 140000 | 36000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| ✓516 | 50000 | 9000 | 0.16 | 1.5 | No | --- | --- | --- | --- | MEA | F | (Acid gas treatment and flare by others) | | | | | | | --- |
| 517 | 33000 | 15100 | -0- | -0- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 518 | 46000 | 27000 | 0.03 | 35.6 | No | --- | --- | --- | --- | HPCA | (Acid gas recompressed separately for reinjection) | | | | | | | | --- |
| 519 | 40000 | 9800 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 520 | 100a | 414 | 0.11 | 10.86 | Yes | DEA | 414 | 0.0288 | 0.0004- | DEA | F60 | --- | --- | --- | --- | --- | --- | --- | --- |
| 521 | 5a | (Not in service) | | --- | Yes | DEA | (Not in service) | | | DEA | V | --- | --- | --- | --- | --- | --- | --- | --- |
| 522 | 40a | (Not in service) | | --- | Yes | DEA | (Not in service) | | | DEA | I25 | --- | --- | --- | --- | --- | --- | --- | --- |
| Note a: Sweetening plant. GPM rated - Flow capacity fluctuates depending upon Acid Gas Concentration in Gas Stream. | | | | | | | | | | | | | | | | | | | |
| 523 | 10000 | 2000 | -0- | 1.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 524 | 5000 | 1500 | 3.0 | 3.0 | No | --- | --- | --- | --- | MEA | F60 | --- | --- | --- | --- | --- | --- | --- | --- |
| 525 | 5000 | 1700 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 526 | 145000 | 135000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 527 | 290000 | 104000 | -0- | 0.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 528 | 220000 | 146000 | -0- | 0.4 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 529 | 120000 | 105000 | -0- | 1.4 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 530 | 25000 | 10000 | 0.003 | 1.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 531 | 150000 | 126000 | -0- | 3.8 | No | --- | --- | --- | --- | MEAI | I50 | --- | --- | --- | --- | --- | --- | --- | --- |
| 532 | 1260000 | 684000 | -0- | 1.1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 533 | 40000 | 27000 | --- | 0.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 534 | 40000 | 31000 | --- | 0.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 535 | [illegible] | 56000 | --- | 0.6 | No | --- | ---- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
|  |  |  |  |  |  |  |  |  |  |  |  | Sulfur Recovery Plant |  |  |  |  |  |  |  |
|  |  |  | Total Plant System |  | Field Gas Sweeting |  |  |  |  |  |  |  |  |  |  |  | Tail Gas Treating |  |  |
|  |  |  |  |  |  |  |  | MOL % $H_2S$ |  |  |  |  |  |  |  |  |  | Exit |  |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Proc- ess Used | Vol (MCFD) | Before | After | Sweet- ening Proc- ess | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Sulf. Equiv #/Day | Sulf. Disp. |
| 536 | 22000 | 21000 | --- | 0.4 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 537 | 26000 | 17000 | 3.0 | 7.0 | Yes | MEA | 17000 | 3.0 | --- | MEA | SR | 2StCl | 23 | 17 | 15 | 4480 | --- | --- | SalTr |
| 538 | 43000 | 58000 | --- | 0.8 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 539 | 250000 | 50000 | --- | 0.1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 540 | 2650000 | 1164000 | --- | 0.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 541 | 35000 | 6000 | --- | 2.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 542 | 15000 | 7000 | --- | 0.2 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| ✓543 | 180000 | 39000 | --- | --- | Yes | MEA | 3700 | 0.2 | --- | MEA | F74 | --- | --- | --- | --- | --- | --- | --- | --- |
| 544 | 235000 | 149000 | --- | 0.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 545 | 100000 | 61000 | --- | 0.02 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 546 | 65000 | 59000 | 1.28 | 1.62 | No | --- | --- | --- | --- | MEA | SR | 2StCl | 23 | 23 | 21.5 | 3360 | --- | --- | SalTr |
| 547 | 33000 | 9000 | Nil | 0.402 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 548 | 5000 | Shutdown | Nil | 2.29 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 549 | 23000 | 23000 | Nil | 0.877 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 550 | 75000 | 16800 | Nil | 0.82 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 551 | 100000 | 18300 | Nil | 0.374 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 552 | 4000 | Shutdown | 0.0032 | 0.88 | No | --- | --- | --- | --- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| ✓553 | 60000 | 7800 | 0.024 | 0.52 | No | --- | --- | --- | --- | DGA | F65 | --- | --- | --- | --- | --- | --- | --- | --- |
| ✓554 | --- | --- | --- | --- | Yes | DGA | 68700 | 0.50 | --- | DGA | I175 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 90,000 MCFD. Inlet MOL % $CO_2$ = 4.80) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 555 | 6000 | 2800 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 555A |  |  | 0.0327 | 0.25 | No | --- | --- | --- | --- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 555B |  |  | -0- | 0.41 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 556 | 42000 | 7600 | 0.0093 | 2.00 | No | --- | --- | --- | --- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 557 | 20000 | 1000 | 5.25 | 4.11 | No | --- | --- | --- | --- | MEA | SR | 3StCl | 4.5 | 3.44 | 3.10 | 762 | I86 | 762 | Sales |
| 558 | --- | --- | --- | --- | Yes | MEA | 9300 | 0.0018 | --- | MEA | F35 | --- | --- | --- | --- | --- | --- | --- |  |
| (Treating Plant. Capacity = 25,000 MCFD. Inlet MOL % $CO_2$ = 6.0) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 559 | 31300 | 11600 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 559A |  |  | -0- | 0.06 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 559B |  |  | 0.1179 | 1.31 | No | --- | --- | --- | --- | MEA | F25 | --- | --- | --- | --- | --- | --- | --- | --- |
| 559C |  |  | 0.0165 | 1.39 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 559D |  |  | 0.1198 | 1.04 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 560 | 25000 | 19300 | -0- | 0.63 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 561 | 10000 | 1300 | -0- | 0.31 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | Total Plant System |  | Field Gas Sweeting |  |  |  |  | Sweet-ening Proc-ess | Acid Gas Disp | Sulfur Recovery Plant |  |  |  |  | Tail Gas Treating |  |  |
|  |  |  | H$_2$S (MOL%) | CO$_2$ (MOL %) | Yes or No | Proc-ess Used | Vol (MCFD) | MOL % H$_2$S Before | MOL % H$_2$S After |  |  | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 562 | --- | --- | --- | --- | Yes | IrSp | 6000 | 0.004 | Nil | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 10,000 MCFD. Inlet MOL % CO$_2$ = 0) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 563 | --- | --- | --- | --- | Yes | Amine | 6000 | 0.02 | Nil | Amine | F35 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 10,000 MCFD. Inlet MOL % CO$_2$ = 1.0) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 564 | --- | --- | --- | --- | Yes | Amine | 3000 | 0.39 | Nil | Amine | F52 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 5000 MCFD. Inlet MOL % CO$_2$ = 5.42) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 565 | --- | --- | --- | --- | Yes | IrSp | 8000 | 1.0 | Nil | IrSp | None | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 30,000 MCFD. Inlet MOL % CO$_2$ = 0) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 566 | --- | --- | --- | --- | Yes | Amine | 4000 | 3.6 | Nil | Amine | SR (Also F35) | 3StCl | 20 | 5.0 | 4.5 | 1120 | --- | --- | Sales |
| (Treating plant. Capacity = 10,000. Inlet MOL % CO$_2$ = 3.67) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 567 | --- | --- | --- | --- | Yes | IrSp | 5000 | 0.8 | Nil | IrSp | None | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 10,000 MCFD. Inlet MOL % CO$_2$ = 0) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 568 | --- | --- | --- | --- | Yes | Amine | 30000 | 0.21 | Nil | Amine | F120 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 40,000 MCFD. Inlet MOL % CO$_2$ = 2.3) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 569 | --- | --- | --- | --- | Yes | Amine | 30000 | 0.21 | Nil | Amine | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 225,000 MCFD. Inlet MOL % CO$_2$ = 2.3) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 570 | --- | --- | --- | --- | Yes | Amine | 35000 | 0.3 | Nil | Amine | V75 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 35,000 MCFD. Inlet MOL % CO$_2$ = 9.0) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 571 | --- | --- | --- | --- | Yes | IrSp | 14000 | 1.75 | Nil | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 30,000 MCFD. Inlet MOL % CO$_2$ = 0) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 572 | 6000 | 2500 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 573 | 5000 | 2000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 574 | 4000 | 1500 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | Total Plant System | | Field Gas Sweeting | | | | | | | Sulfur Recovery Plant | | | | | Tail Gas Treating | | |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After | Sweetening Process | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 575 | 150000 | 65000 | 0.39 | 0.27 | No | --- | --- | --- | --- | MEA | F160 | --- | --- | --- | --- | --- | --- | --- | --- |
| 576 | 120000 | 90000 | 0.15 | 0.80 | No | --- | --- | --- | --- | MEA | F115 | --- | --- | --- | --- | --- | --- | --- | --- |
| 577 | 16000 | 10000 | 0.40 | 0.62 | No | (Gas processed for liq. only and full sour gas stream returned to a gas company.) | | | | | | | | | | | | | |
| 578 | 180000 | | | | | | | | | | | | | | | | | | |
| 578A | | 141000 | 0.03 | 0.55 | No | --- | --- | --- | --- | MEA | F210 | --- | --- | --- | --- | --- | --- | --- | --- |
| | | | | | | | | | | | (Also I210) | | | | | | | | |
| 578B | | 9000 | 0.0015 | Trace | Yes | MEA | 9000 | 1.0 | 0.0015 | DEA | F60 | (SULFUR RECOVERED BY OTHERS | | | | | --- | --- | --- |
| 579 | 210000 | --- | --- | --- | --- | --- | --- | --- | --- | MEA | F60 | --- | --- | --- | --- | --- | --- | --- | --- |
| 579A | | 38000 | Trace | 0.07 | No | | | | | | | | | | | | | | |
| 579B | | 115000 | 0.07 | --- | No | | | | | | | | | | | | | | |
| 579C | | 41000 | Trace | --- | No | | | | | | | | | | | | | | |
| 580 | 60000 | | | | | | | | | | | | | | | | | | |
| 580A | | 47000 | Trace | 0.19 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 580B | | 3500 | 0.016 | 0.24 | No | --- | --- | --- | --- | MEA | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 581 | --- | --- | --- | --- | Yes | DEA | 1500 | 1.0 | 0.001 | DEA | F70 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. Capacity = 6000 MCFD. No inlet $CO_2$ indicated.) | | | | | | | | | | | | | | | | | | | |
| 582 | 100000 | 77600 | -0- | 0.40 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 583 | 25000 | 18300 | 0.1 | 0.23 | No | --- | --- | --- | --- | --- | I | --- | --- | --- | --- | --- | --- | --- | --- |
| | | | | | | | | | | | (Tip Incinerator) | | | | | | | | |
| 584 | 11600 | 7500 | -0- | 0.09 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 585 | 30000 | 6700 | 30.0 | 6.5 | No | --- | --- | --- | --- | Sulf | SR | 3StCl | 159 | 77.2 | 76.0 | 2688 | --- | --- | Sold |
| 586 | 38000 | 4600 | -0- | 0.13 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 587 | 50000 | 14000 | -0- | 1.02 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 588 | 20000 | 4300 | -0- | 0.46 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
|  |  |  | Total Plant System |  | Field Gas Sweeting |  |  |  |  |  |  | Sulfur Recovery Plant |  |  |  |  |  |  |  |
|  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  | Tail Gas Treating |  |  |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | H₂S (MOL%) | CO₂ (MOL %) | Yes or No | Proc-ess Used | Vol (MCFD) | MOL % H₂S Before | MOL % H₂S After | Sweet-ening Proc-ess | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 589 | 75000 | 22570 | -0- | 0.87 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 590 | 4000 | 2510 | -0- | 0.43 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 591 | 50000 | 3720 | -0- | 0.14 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 592 | 85000 | 62000 | -0- | 0.15 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 593 | 65000 | 7380 | --- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 594 | 12400 | 1600 | -0- | 0.40 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 595 | 2500 | 1340 | 2.2 | 7.53 | No | --- | --- | --- | --- | MEA | I | --- | --- | --- | --- | --- | --- | --- | --- |
| 596 | 40000 | 30650 | 2.0 | 4.9 | No | --- | --- | --- | --- | MEAa | SR | 2StCl | 10 | 13.5 | 7.56 | 134406 | --- | --- | Sold |

(Notes: a. MEA treats 50%, reinject 50%.
b. Col. 17 & 18 values are operating averages--annual averages are 5.9 and 17024.)

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| 597 | 15000 | 1560 | -0- | 0.95 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 598 | 5000 | 1600 | --- | 0.79 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 599 | 20000 | 8000 | 0.03 | 0.93 | No | --- | --- | --- | --- | MEA | F40 | --- | --- | --- | --- | --- | --- | --- | --- |
| 599A | 9000 | -0- | --- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 6000 | 150000 | 16000 | -0- | 0.8 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

(Processes only offshore gas.)

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| 601 | 10000 | 2850 | 0.45 | 0.8 | No | --- | --- | --- | --- | MEA | F30 | --- | --- | --- | --- | --- | --- | --- | --- |
| 602 | 26000 | 6500 | --- | 0.13 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 603 | 5000 | 4000 | -0- | 0.13 | No | --- | --- | --- | --- | (Sulfur treating by others--see Plant No. 566) |  |  |  |  |  |  |  |  |  |
| 604 | 35000 | 40000 | 0.0004 | 3.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 605 | 35000 | 30000 | 0.0004 | 3.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 606 | 15000 | 10000 | 0.0004 | 3.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 607 | 240000 | 255000 | -0- | 1.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 608 | 12000 | 9000 | -0- | 2.00 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 609 | 135000 | 105000 | -0- | 1.47 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
|  |  |  | Total Plant System |  | Field Gas Sweeting |  |  |  |  |  |  | Sulfur Recovery Plant |  |  |  |  |  |  |  |
|  |  |  |  |  |  |  |  | MOL % $H_2S$ |  |  |  |  |  |  |  |  | Tail Gas Treating |  |  |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | Before | After | Sweetening Process | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 610 | 30000 | 30000 | Nil | 4.52 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 611 | 50000 | 35000 | 0.11 | 0.80 | No | --- | --- | --- | --- | Amine | F60 | --- | --- | --- | --- | --- | --- | --- | --- |
| 612 | 25000 | 25000 | 0.13 | 0.36 | No | --- | --- | --- | --- | Amine | F60 | --- | --- | --- | --- | --- | --- | --- | --- |
| 613 | 150000 | 125000 | Nil | 1.10 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 614 | 50000 | 50000 | Nil | 0.57 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 615 | 15000 | 6000 | 0.67 | 0.28 | No | --- | --- | --- | --- | Amine | F107 | --- | --- | --- | --- | --- | --- | --- | --- |
| 616 | 50000 | 28512 | 6.46 | 3.92 | No | --- | --- | --- | --- | DGA | SR | 3StCL | 200 | 116.6 | 113.3 | 11872 | --- | --- | Sales |
| 617 | 40000 | 7627 | -0- | 4.58 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 618 | 30000 | 7147 | -0- | 0.50 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 619 | 15000 | 7778 | 27.20 | 4.44 | No | --- | --- | --- | --- | DGA | SR | 2StCl | 135 | 71.6 | 68.4 | 7168 | --- | --- | Sales |
| 620 | 196000 | 185000 | 0.009 | 0.74 | No | --- | --- | --- | --- | Amine | F175 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Note: Inlet not treated. Amine plant treats NGL product containing 1100#/Day $H_2S$ and 64000#/Day $Co_2$.) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 621 | 8000 | 5600 | Nil | 0.5 | No | --- | --- | --- | --- | AmC | V50 | (AmC removes $CO_2$ to meet liquid product spec.) |  |  |  |  |  |  |  |
| 622 | 4500 | 3100 | Nil | 0.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 623 | 15000 | 6500 | Nil | 0.4 | No | --- | --- | --- | --- | AmC | V50 | (AmC removes $CO_2$ to meet liquid product spec.) |  |  |  |  |  |  |  |
| 624 | 10000 | 4000 | Nil | 0.4 | No | --- | --- | --- | --- | AmC | V40 | (AmC removes $CO_2$ to meet liquid product spec.) |  |  |  |  |  |  |  |
| 625 | 4000 | 2000 | Nil | Nil | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 626 | 18000 | 16000 | 0.2 | 0.2 | No | --- | --- | --- | --- | Amine | F75 | --- | --- | --- | --- | --- | --- | --- | --- |
| 627 | 2000 | 600 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 628 | 21000 | 7500 | 0.001- | 0.38 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 629 | 25000 | 23000 | -0- | 0.61 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 630 | 22245 | 1800 | 7.50 | 7.50 | No | --- | --- | --- | --- | MEA | SR<br>(Also 170) | Claus | 75 | 5.1 | 4.5 | 1344 | --- | --- | Truck |
| 631 | 25000 | 5829 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 632 | 33000 | 22180 | -0- | 0.13 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 633 | 100000 | 37487 | 2.15 | 6.00 | No | --- | --- | --- | --- | DEA | SR<br>(Also I50) | Claus | 36 | 30.3 | 29.0 | 2912 | --- | --- | Truck |
| 634 | 25000 | 9962 | 0.02 | 1.24 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | | | | | | | | Sulfur Recovery Plant | | | | | | | |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | Total Plant System | | Field Gas Sweeting | | | | | Sweet-ening Proc-ess | Acid Gas Disp | | | | | | Tail Gas Treating | | |
| | | | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Proc-ess Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After | | | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 635 | 5000 | 1795 | 0.06 | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 636 | 9000 | 8345 | -0- | 0.28 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 637 | 45000 | 22718 | 0.21 | 0.25 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 638 | 10000 | --- | --- | --- | No | | | | | | | | | | | | | | |
| 638A | | 500 | 6.0 | 0.53 | --- | --- | --- | --- | --- | Amine | F40 | --- | --- | --- | --- | --- | --- | --- | --- |
| 638B | | 9500 | -0- | 0.53 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 639 | 10000 | 2153 | -0- | 0.28 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 640 | 155000 | 78666 | 1.05 | 1.43 | No | --- | --- | --- | --- | MEA | SR (Also I60) | Claus | 45 | 31 | 27.0 | 8960 | --- | --- | Truck |
| 641 | 14750 | 9468 | 0.3 max | 0.06 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 642 | 7000 | 4000 | -0- | 0.34 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 643 | 5000 | 1693 | 0.01 | 1.7 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 644 | 1000 | 699 | 2.44 | 8.34 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 645 | 5000 | 1616 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 646 | 4000 | 4184 | -0- | 0.55 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 647 | 105000 | 62063 | 1.8 | 2.5 | No | --- | --- | --- | --- | DEA | SR (Also I75) | Claus | 75 | 42 | 37.0 | 11200 | --- | --- | Truck |
| 648 | 15000 | 11477 | -0- | 0.41 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 649 | 11000 | 3000 | 0.4 | 5.6 | No | --- | --- | --- | --- | MEA | F80 | --- | --- | --- | --- | --- | --- | --- | --- |
| 650 | 180000 | 38700 | --- | 1.5 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 651 | 60000 | --- | --- | --- | --- | --- | --- | --- | --- | DEA | SR | 3StCl | 308 | 177 | 174 | 6720 | --- | 6720 | Sales |
| 651A | | 20000 | 17 | 3.0 | No | | | | | | | | | | | | | | |
| 651B | | 32000 | 8 | 2.0 | No | | | | | | | | | | | | | | |
| 652 | 51000 | 51000 | --- | 0.46 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 652A | 31000 | 31000 | --- | --- | | | | | | | | | | | | | | | |
| 653 | 10000 | 7500 | 2.60 | 5.90 | No | --- | --- | --- | --- | DGA | F60 | --- | --- | --- | --- | --- | --- | --- | --- |
| 654 | 36000 | 22500 | 0.27 | 1.7 | No | --- | --- | --- | --- | MEA | F120 | --- | --- | --- | --- | --- | --- | --- | Sales |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
|  |  |  |  |  |  |  |  |  |  |  |  | Sulfur Recovery Plant |  |  |  |  |  |  |  |
|  |  |  | Total Plant System |  | Field Gas Sweeting |  |  |  |  |  |  |  |  | (A) | (B) | (A)-(B) | Tail Gas Treating |  |  |
|  |  |  |  |  |  |  |  | MOL % $H_2S$ |  |  |  |  |  |  |  |  |  |  |  |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Proc-ess Used | Vol (MCFD) | Before | After | Sweet-ening Proc-ess | Acid Gas Disp | Type | Design Cap LT/Day | Avail. Sulf LT/Day | Corres Meas. Rec. LT/Day | Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 655 | 90000 | 20900 | --- | 0.16 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 656 | 45000 | 10400 | 3.8 | 2.8 | No | --- | --- | --- | --- | MEA | SR | 2StCl | 22.0 | 14.5 | 14.4 | 224 | CBA | --- | Sales |
|  | (Compression capacity. Process Cry capacity is 12000 MCFD) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| ✓657 | 12000 | 5900 | 0.07 | 0.6 | No | --- | --- | --- | --- | MEA | F47 | --- | --- | --- | --- | --- | --- | --- | --- |
| 658 | 37000 | 19900 | 4.7 | 2.5 | No | --- | --- | --- | --- | MEA | SR | 2StCl | 30.0 | 25.6 | 24.21 | 3110 | --- | --- | Sales |
| 659 | 570000 | 205000 | --- | 1.3 | No | --- | --- | --- | --- | DEAC | V80 | --- | --- | --- | --- | --- | --- | --- | --- |
| 660 | 72000 | 34300 | 3.5 | 10.0 | No | --- | --- | --- | --- | MEA | SR | 2StCl | 42.0 | 30.0 | 27.6 | 5380 | --- | 5380 | Sales |
| 661 | 10000 | 10150 | 0.09 | 1.2 | No | --- | --- | --- | --- | MEA | SR | 2StCl | 6.0 | 3.5 | 3.1 | .900 | --- | 900 | Sales |
| 662 | 360000 | 36300 | --- | 3.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 663 | 50000 | 11100 | 5.4 | 7.7 | No | --- | --- | --- | --- | MEA | SR | 2StCl | 30.0 | 28.0 | 26.0 | 4480 | --- | 4480 | Sales |
| 664 | 20000 | 8700 | 0.0005 | 2.25 | Yes | Amine | 8700 | -0.0005 | Nil | Amine | F30 | --- | --- | --- | --- | --- | --- | --- | --- |
| 665 | 5000 | -0- | -0- | --- | Yes | Amine | -0- | -0- | -0- | Amine | F50 | --- | --- | --- | --- | --- | --- | --- | --- |
| 666 | 20000 | 7000 | --- | --- | Yes | IrSp | 7000 | --- | -0- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 667 | 5000 | -0- | -0- | --- | Yes | IrSp | -0- | --- | --- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 668 | (60 GPM) | -0- | -0- | --- | Yes | AIrSp | -0- | --- | --- | AIrSp | I100 | --- | --- | --- | --- | --- | --- | --- | --- |
| 669 | 20000 | 16000 | -0- | 0.20 | No | --- | --- | --- | --- | DEA | V30 | --- | --- | --- | --- | --- | --- | --- | --- |
| 670 | 11000 | 14000 | 0.30 | 0.30 | No | --- | --- | --- | --- | MEA | F100<br>(Also I2.25) | --- | --- | --- | --- | --- | --- | --- | --- |
| 671 | 24000 | 12000 | 0.10 | 0.22 | No | --- | --- | --- | --- | DEA | F67 | --- | --- | --- | --- | --- | --- | --- | --- |
| 672 | 15000 | 9000 | 0.55 | 8.15 | No | --- | --- | --- | --- | DEA | F100 | --- | --- | --- | --- | --- | --- | --- | --- |
| 673 | 33500 | 29000 | 1.00 | 3.31 | No | --- | --- | --- | --- | DEA | F116<br>(Also I2.25) | --- | --- | --- | --- | --- | --- | --- | --- |
| 674 | 6500 | 3000 | 0.07 | 0.26 | No | --- | --- | --- | --- | DEA | F40 | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Plant or Stream No. | Current Capacity (MCFD) | Current Throughput (MCFD) | Total Plant System | | Field Gas Sweeting | | | | | Sweetening Process | Acid Gas Disp | Sulfur Recovery Plant | | | | | | | |
| | | | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % $H_2S$ | | | | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Tail Gas Treating | | |
| | | | | | | | | Before | After | | | | | | | | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 675 | 70000 | 70352 | 5.52 | 6.60 | No | --- | --- | --- | --- | Sulf | SR | Claus | 187 | 122 | 118 | 8960 | Incin | --- | Sales |
| 676 | 6000 | 1405 | --- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 677 | 70000 | 38400 | 31.85 | 5.54 | No | --- | --- | --- | --- | Sulf | SR | Claus | 842 | 385 | 384 | 1656a<br>(Note a: EPA Permit) | Scot | 6000a | Sales |
| 678 | --- | --- | --- | --- | Yes | Sulf | 7000 | 0.003 | 0.0004 | Sulf | V60 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating Plant. | Capacity = 20,000 MCFD. | | | Inlet MOL % $CO_2$ = 15.16) | | | | | | | | | | | | | | | |
| 679 | 285000 | 137396 | --- | 3.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 680 | 10000 | 6822 | --- | 0.62 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 681 | --- | --- | --- | --- | Yes | Sulf | 10000 | 0.02 | 0.0004 | Sulf | I130 | --- | --- | --- | --- | --- | --- | --- | --- |
| (Treating plant. | Capacity = 50,000 MCFD. | | | Inlet MOL % $CO_2$ = 8.16) | | | | | | | | | | | | | | | |
| 682 | 68500 | 20528 | 1.16 | 5.7 | No | --- | --- | --- | --- | Sulf | SR | Claus | 32 | 7.5 | 7 | 1120 | Incin | --- | Sales |
| 683 | 35000 | 29944 | --- | 0.11 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 684 | 40000 | 33000 | 0.55 | 0.77 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 685 | 175000 | 120474 | 0.25 | 5.71 | No | --- | --- | --- | --- | (RESIDUE GAS TREATED BY OTHERS) | | | | | | | | | |
| 686 | 89800 | 50000 | 0.0006 | 2.2 | No | --- | --- | --- | --- | F197 | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 687 | 24400 | 5700 | -0- | 0.31 | Yes | Amine<br>(At Plant inlet) | 5700 | 0.1 | -0- | Amine | F33 | --- | --- | --- | --- | --- | --- | --- | --- |
| 688 | 43200 | 33500 | -0- | 0.8 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
|  |  |  | Total Plant System |  | Field Gas Sweeting |  |  |  |  |  |  | Sulfur Recovery Plant |  |  |  |  | Tail Gas Treating |  |  |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Proc-ess Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After | Sweet-ening Proc-ess | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 689 | 22000 | 8900 | 0.008 | 0.75 | No | --- | --- | --- | --- | IrSp | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 690 | 12500 | 1900 | 0.008 | 0.45 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 691 | 20000 | 18000 | -0- | 1.034 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 692 | 80000 | 51500 | 1.08 | 27.25 | No | --- | --- | --- | --- | DPrC | F50 | --- | ---- | ---- | --- | --- | --- | --- | --- |
| (Note Re Col. 12: Acid gas removed in fuel stream only. Residue gas re-injected at 28% $CO_2$ and .94% $H_2S$ to reservoir.) |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 693 | 38000 | 8900 | 0.024 | 0.50 | Yes | IrSp | 10000 | 0.024 |  | 0.0004 |  | --- | --- | --- | --- | --- | --- | --- | --- |
| 694 | 80000 | 50000 | -0- | 0.3 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 695 | 40000 | 15368 | --- | 0.26 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 696 | 150000 | 117000 | 0.0005- | 0.10- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 697 | 70000 | 70000 | 0.0004- | 0.1 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 698 | 30000 | 15000 | 0.0032 | 65 | No | --- | --- | --- | --- | Sele | Sold | --- | --- | --- | --- | --- | --- | --- | --- |
| 699 | 170000 | 120000 | 0.0004- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

## API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | Total Plant System | | Field Gas Sweeting | | | | | Sweetening Process | Acid Gas Disp | Sulfur Recovery Plant | | | | | Tail Gas Treating | | |
| | | | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | MOL % $H_2S$ Before | MOL % $H_2S$ After | | | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 700 | 9000 | 2000 | --- | Trace | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 701 | 60000 | 50000 | 3.8 | 17.5 | No | --- | --- | --- | --- | Bfd | SR | 2StCl | 82 | 82 | 68 | 31360 | ATS | 200 | Sales |
| 702 | 500 | 190 | -0- | 0.85 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 703 | 2200 | 1100 | -0- | 0.70 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 704 | 80000 | 65000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 705 | 6000 | 3000 | -0- | -0- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 706 | 250000 | 206000 | -0- | 0.27 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 707 | 30000 | 30000 | -0- | 0.55 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 708 | 6000 | :2561 | -0- | 3.177 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 709 | 31000 | 7500 | 0.0004- | 0.82 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 710 | 30000 | --- | --- | --- | --- | --- | --- | --- | --- | MEA | F60 | --- | --- | --- | --- | --- | --- | --- | --- |
| 710A | | 8000 | 0.0004- | 0.19 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 710B | | 1500 | 1.48 | 1.84 | Yes | MEA | 1500 | 2.2 | 0.0004 | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 711 | 10000 | 10500 | Nil | 1.80 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 712 | 12000 | 4700 | Nil | 1.20 | No | | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 713 | 7200 | 3200 | 14.9 | 1.59 | No | --- | --- | --- | --- | DGA | SR | 3StCl | 15 | 13 | 11.2 | 4032 | GLCU | $SO_2$ (4.4 | Vent MOL/HR) |
| 714 | 2000 | Startup | 1.37 | 1.27 | No | --- | --- | --- | --- | DGA | F100 | --- | --- | --- | --- | --- | --- | --- | --- |
| 715 | 75000 | 50000 | -0- | 1.338 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 716 | 50000 | 14400 | -0- | 0.75 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 717 | 20000 | 6000 | -0- | 1.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 718 | 40000 | 26400 | -0- | 1.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 719 | 10000 | 3000 | -0- | 1.2 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 720 | --- | --- | | | | | | | | (All acid gas injected back into producing reservoir.) | | | | | | | | | |
| 720A | 55000 | 50000 | 1.5 | 29.8 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 720B | 8000 | 5000 | 0.5 | 75.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 720C | 2000 | 2000 | 0.5 | 34.9 | No | --- | --- | --- | --- | DEA | (Injected back into producing reservoir) | | | | | | | | |
| 721 | 27000 | 15000 | -0- | 2.0 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 722 | 4000 | 450 | -0- | 0.8 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 723 | 35000 | 40000 | Nil | 0.2 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

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API GAS PLANT SURVEY

| 1 | 2 | 3 | 4 | 5 | 6/7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
|  |  |  |  |  |  |  |  |  |  |  |  | Sulfur Recovery Plant |  |  |  |  |  |  |  |
|  |  |  | Total Plant System |  | Field Gas Sweeting |  |  |  |  |  |  |  |  |  |  |  | Tail Gas Treating |  |  |
|  |  |  |  |  |  |  |  | MOL % $H_2S$ |  |  |  |  |  |  |  |  |  |  |  |
| Plant or Stream No. | Current Capacity (MCFD) | Current Through put (MCFD) | $H_2S$ (MOL%) | $CO_2$ (MOL %) | Yes or No | Process Used | Vol (MCFD) | Before | After | Sweetening Process | Acid Gas Disp | Type | Design Cap LT/Day | (A) Avail. Sulf LT/Day | (B) Corres Meas. Rec. LT/Day | (A)-(B) Equiv. Sulf. Diff. #/Day | Type | Exit Sulf. Equiv #/Day | Sulf. Disp. |
| 724 | 4000 | 2500 | 0.35 | 20.6 | No | --- | --- | --- | --- | DEA | V115 | --- | --- | --- | --- | --- | --- | --- | --- |
| 725 | 65000 | 29600 | --- | 1.8 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 726 | 17500 | 5400 | 6.0 | 2.0 | No | --- | --- | --- | --- | MEA | SR | Claus | 60 | 16.1 | 14.5 | 3584 | --- | 3584 | Sales |
| 727 | 17500 | 11700 | 10.0 | 8.0 | No | --- | --- | --- | --- | MEA | SR | Claus | 75 | 53 | 50 | 6720 | --- | 6720 | Sales |
| 728 | 14000 | 14000 | --- | 0.07 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 729 | 100000 |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |  |
| 729A |  | 16000 | --- | --- | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |
| 729B |  | 8000 | --- | --- | Yes | DGA | 12500<br>(Design Basis) | 12 | 0.0004- | DGA | SR | Clau | 70 | ---<br>(Not yet on stream) | --- | --- | CBA | 1252 | Sales |
| 729C |  | 1000 | --- | --- | Yes | DGA | 3000<br>(Design Basis) | 7 | 0.0004- | DGA | I | --- | --- | --- | --- | --- | --- | --- | --- |
| 730 | 6000 | 3000 | 1.60 | 6.50 | No | --- | --- | --- | --- | MEA | F60 | --- | --- | --- | --- | --- | --- | --- | --- |
| 731 | 10000 | 2500 | -0- | 0.70 | No | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- |

# SMALL SULFUR RECOVERY UNITS

## ONSHORE SOUR GAS PRODUCTION FACILITIES

APPENDIX H

# SECTION 1

# DESIGN BASIS

## INTRODUCTION

This study was prepared for TRW to assist the U.S. Environmental Protection Agency in promulgating New Source Performance Standards for Onshore Production Facilities. Specifically, this study is to provide data to enable determination of cost-effectiveness for different sizes of small sulfur recovery units with feed acid gases of low $H_2S$ content.

This report provides investment costs, direct operating cost data, process descriptions, process flow diagrams, and atmospheric emissions for the following 12 cases:

| Case No. | Sulfur Feed Rate (LT/D) | Mol % $H_2S$ in Acid Gas | Sulfur Recovery Process |
|---|---|---|---|
| 1 | 0.5 | 2.0 | Selectox Two-Stage |
| 2 | 0.5 | 5.0 | Recycle Selectox Two-Stage |
| 3 | 0.5 | 12.5 | Recycle Selectox Two-Stage |
| 4 | 2.0 | 2.0 | Selectox Two-Stage |
| 5 | 2.0 | 5.0 | Recycle Selectox Two-Stage |
| 6 | 2.0 | 12.5 | Recycle Selectox Two-Stage |
| 7 | 2.0 | 2.0 | Selectox Three-Stage |
| 8 | 2.0 | 5.0 | Recycle Selectox Three-Stage |
| 9 | 2.0 | 12.5 | Recycle Selectox Three-Stage |
| 10 | 5.0 | 2.0 | Selectox Three-Stage |
| 11 | 5.0 | 5.0 | Recycle Selectox Three-Stage |
| 12 | 5.0 | 12.5 | Recycle Selectox Three-Stage |

For explanations of terminology and abbreviations used, refer to Definitions in this section.

## DESIGN CRITERIA AND ASSUMPTIONS

Listed below are assumptions used for design criteria and cost estimating for this study:

(1) Barometric pressure is 14.7 psia.

(2) Acid gases are available at 120°F and 24.7 psia, saturated with water, and contain 0.5 vol% methane (wet basis) maximum.

(3) Investment costs in Table 2-2 are based on February 1983 Gulf Coast prices and include the complete plants through the catalytic incinerator and 100-ft-high stack.

(4) The plants use a heating medium for the preheaters and reheaters and a cooling medium for the sulfur condensers. Excess heat is removed from the cooling medium by a cooler, but can also be recovered by circulating it through a reboiler on the regenerator in the adjacent amine unit. For these small plants, such a reboiler has not been included as the economics may be marginal.

## DEFINITIONS

The following list provides definitions, terminology, and abbreviations used in this report:

Acid Gas: The gas containing $H_2S$ (and $CO_2$) resulting from treating of sour natural gas. Also contains small amounts of hydrocarbons and water.

Catalytic Incineration: Process by which sulfur compounds are catalytically oxidized to $SO_2$.

Claus: The name of the gas phase reaction where 2 moles of $H_2S$ combine with 1 mole of $SO_2$ to form 3 moles of sulfur and 2 moles of water.

End-of-Run: After a certain period of operation, the activity of a catalyst declines to a point where it is economical to shut down and replace it with fresh catalyst. End-of-run is defined as that point when the percent recovery has declined to a specified value. For this study, average lives of 5 years for the Selectox catalyst and 3 years for the alumina catalyst have been assumed. The $SO_2$ emissions in the tables in this report are based on end-of-run recoveries. The LT/D of sulfur recovered in Table 4-2 are average during run, calculated by averaging the start-of-run and end-of-run recoveries.

| Gas Components: | | |
|---|---|---|
| | $H_2S$ | Hydrogen Sulfide |
| | $SO_2$ | Sulfur Dioxide |
| | $O_2$ | Oxygen |

| | |
|---|---|
| $N_2$ | Nitrogen |
| $CO_2$ | Carbon Dioxide |
| $H_2O$ | Water |
| $CH_4$ | Methane |

**Lb mol/hr**: The pounds per hour divided by the molecular weight of the component.

**LT/D**: Long tons (2,240 pounds) per day.

**MM**: Millions.

**Selectox**: A catalyst developed by Union Oil Company of California which oxidizes $H_2S$ directly to elemental sulfur and in a temperature range substantially lower than required for uncatalyzed oxidation.

**Sour Gas**: Natural gas containing $H_2S$. When the $H_2S$ and usually the $CO_2$ are absorbed from the hydrocarbons by a suitable solvent in any of the various treating units, and then stripped from the solvent, the stripped gas containing the $H_2S$ and $CO_2$ is called acid gas.

**Tail Gas**: Exit gases from a sulfur recovery unit.

**Vol ppm or ppmv**: The parts per million of a component on a volume basis.

## FEED GAS COMPOSITIONS

The feed gas compositions are provided in Table 1-1. The gases are saturated with water at the conditions stated in Design Criteria and Assumptions. These are representative, although the water and hydrocarbon contents vary from plant to plant depending on design and operation of the upstream facilities.

**Table H-1 - Feed Gas Compositions**
(Table 1-1)

| % $H_2S$ | 2.0 | | | 5.0 | | | 12.5 | | |
|---|---|---|---|---|---|---|---|---|---|
| **Sulfur Input (LT/D)** | 0.5 | 2.0 | 5.0 | 0.5 | 2.0 | 5.0 | 0.5 | 2.0 | 5.0 |
| **Case No.** | 1 | 4,7 | 10 | 2 | 5,8 | 11 | 3 | 6,9 | 12 |
| **Mol/hr** | | | | | | | | | |
| $H_2S$ | 1.46 | 5.82 | 14.55 | 1.46 | 5.82 | 14.55 | 1.46 | 5.82 | 14.55 |
| $CO_2$ | 65.71 | 262.85 | 657.13 | 25.41 | 101.65 | 254.12 | 9.29 | 37.16 | 92.91 |
| $CH_4$ | 0.36 | 1.46 | 3.64 | 0.15 | 0.58 | 1.46 | 0.06 | 0.23 | 0.58 |
| $H_2O$ | 5.24 | 20.96 | 52.40 | 2.10 | 8.38 | 20.95 | 0.84 | 3.35 | 8.38 |
| **Total** | 72.77 | 291.09 | 727.73 | 29.12 | 116.43 | 291.08 | 11.65 | 46.56 | 116.42 |

## SECTION 2

## SUMMARY OF RESULTS

The compositions of the exit gases at end-of-run are given in Table 2-1, and the sulfur recoveries for start-of-run and end-of-run are shown in Table 4-2, Section 4.

Total installed costs are listed in Table 2-2. It is evident from observing the investment costs of the 2 LT/D sulfur input cases that the differential between two-stage and three-stage units becomes smaller as the gas gets richer in $H_2S$.

A discussion of service facilities that might be required is provided in this section.

Information necessary for calculating operating costs is shown in Section 4.

Table H-2 - Composition of Exit Gases (lb mol/hr) End-of-Run
(Table 2-1)

| Case No. | Sulfur Input (LT/D) | % $H_2S$ in Acid Gas | No. of Stages | Type | $SO_2$ | $H_2O$ | $O_2$ | $N_2$ | $CO_2$ | $CH_4$ | Total |
|---|---|---|---|---|---|---|---|---|---|---|---|
| 1 | 0.5 | 2 | 2 | Selectox | 0.37 | 6.81 | 0.17 | 4.61 | 65.41 | 0.36 | 77.73 |
| 2 | 0.5 | 5 | 2 | Recycle Selectox | 0.27 | 3.63 | 0.11 | 4.00 | 25.41 | 0.15 | 33.57 |
| 3 | 0.5 | 12.5 | 2 | Recycle Selectox | 0.21 | 2.39 | 0.08 | 3.68 | 9.29 | 0.06 | 15.71 |
| 4 | 2.0 | 2 | 2 | Selectox | 1.47 | 27.26 | 0.66 | 18.43 | 261.65 | 1.46 | 310.93 |
| 5 | 2.0 | 5 | 2 | Recycle Selectox | 1.07 | 14.53 | 0.45 | 16.01 | 101.65 | 0.58 | 134.29 |
| 6 | 2.0 | 12.5 | 2 | Recycle Selectox | 0.83 | 9.55 | 0.33 | 14.73 | 37.16 | 0.23 | 62.83 |
| 7 | 2.0 | 2 | 3 | Selectox | 1.34 | 27.24 | 0.59 | 17.63 | 261.65 | 1.46 | 309.91 |
| 8 | 2.0 | 5 | 3 | Recycle Selectox | 0.93 | 14.60 | 0.37 | 15.19 | 101.65 | 0.58 | 133.32 |
| 9 | 2.0 | 12.5 | 3 | Recycle Selectox | 0.72 | 9.54 | 0.27 | 14.02 | 37.17 | 0.23 | 61.95 |
| 10 | 5.0 | 2 | 3 | Selectox | 3.36 | 68.09 | 1.48 | 44.08 | 654.13 | 3.64 | 774.78 |
| 11 | 5.0 | 5 | 3 | Recycle Selectox | 2.33 | 36.49 | 0.93 | 37.98 | 254.12 | 1.46 | 333.31 |
| 12 | 5.0 | 12.5 | 3 | Recycle Selectox | 1.80 | 23.84 | 0.69 | 35.06 | 92.92 | 0.58 | 154.89 |

## Table H-3 - Total Installed Costs
(Table 2-2)

| Case No. | Sulfur Input (LT/D) | % $H_2S$ in Acid Gas | No. of Stages | Type | $MM |
|---|---|---|---|---|---|
| 1 | 0.5 | 2 | 2 | Selectox | 1.71 |
| 2 | 0.5 | 5 | 2 | Recycle Selectox | 1.60 |
| 3 | 0.5 | 12.5 | 2 | Recycle Selectox | 1.53 |
| 4 | 2.0 | 2 | 2 | Selectox | 2.17 |
| 5 | 2.0 | 5 | 2 | Recycle Selectox | 1.95 |
| 6 | 2.0 | 12.5 | 2 | Recycle Selectox | 1.80 |
| 7 | 2.0 | 2 | 3 | Selectox | 2.51 |
| 8 | 2.0 | 5 | 3 | Recycle Selectox | 2.20 |
| 9 | 2.0 | 12.5 | 3 | Recycle Selectox | 1.98 |
| 10 | 5.0 | 2 | 3 | Selectox | 3.32 |
| 11 | 5.0 | 5 | 3 | Recycle Selectox | 2.69 |
| 12 | 5.0 | 12.5 | 3 | Recycle Selectox | 2.36 |

**Table** H-4 **- Cost of Initial Charge of Catalysts**
(Table 2-3)

| Case No. | Sulfur Input (LT/D) | % $H_2S$ in Acid Gas | No. of Stages | Type | Cost ($) |
|---|---|---|---|---|---|
| 1 | 0.5 | 2 | 2 | Selectox | 8,800 |
| 2 | 0.5 | 5 | 2 | Recycle Selectox | 4,300 |
| 3 | 0.5 | 12.5 | 2 | Recycle Selectox | 2,600 |
| 4 | 2.0 | 2 | 2 | Selectox | 35,200 |
| 5 | 2.0 | 5 | 2 | Recycle Selectox | 17,100 |
| 6 | 2.0 | 12.5 | 2 | Recycle Selectox | 10,200 |
| 7 | 2.0 | 2 | 3 | Selectox | 37,000 |
| 8 | 2.0 | 5 | 3 | Recycle Selectox | 17,800 |
| 9 | 2.0 | 12.5 | 3 | Recycle Selectox | 10,500 |
| 10 | 5.0 | 2 | 3 | Selectox | 92,400 |
| 11 | 5.0 | 5 | 3 | Recycle Selectox | 44,400 |
| 12 | 5.0 | 12.5 | 3 | Recycle Selectox | 26,000 |

## SERVICE FACILITIES

With regard to offsites, only two utilities are required: (1) natural gas, which is readily available, and (2) electrical power, which is available at most locations.

For very remote locations where electrical power is not available, gas engine generators can be used. The small sulfur recovery plants use only 50 to 100 kW, and the entire adjacent facilities would use from 500 to 1,000 kW. The investment cost and fuel for gas engine generators of this size are as follows:

| Item | Generator Size | |
|---|---|---|
| | 500 kW | 1,000 kW |
| Investment Cost ($) | 430,000 | 830,000 |
| Natural Gas (MMBtu/hr) | 5.5 | 11.0 |

The cost of the natural gas, assuming $5.00/MMBtu, is equivalent to $0.055/kWh.

Since power generation would have to be installed for the rest of the facilities, the incremental cost over purchased power for the small amount of power required for the sulfur recovery unit would not be a significant factor in the total cost of sulfur recovery.

Molten sulfur can be hauled away in tank trucks for sale, or poured into blocks and stored indefinitely to be broken up later and hauled away as solid sulfur. Costs in this study include a simple sulfur loading rack but no other transportation facilities.

## RECOMMENDATIONS AND SUGGESTIONS

While this study was based on very weak gases, from 2% to 12.5% $H_2S$, the Recycle Selectox process works very well on acid gases containing up to 100% $H_2S$. The recycle system merely increases in size, but the cost is still less than Claus-type sulfur recovery units and does not involve boiler feedwater treatment and steam boiler operations. It has the additional advantage of pushbutton operation; in response to a power failure or other shutdown causes, the plant can be stopped and restarted by pushing the necessary buttons. These factors are mentioned to emphasize that the process is practical and economical for gases stronger in $H_2S$ than those covered in this study.

The total installed costs for the sulfur recovery units seem to level out as the plants get smaller.

No suggestions for further study are apparent at this time.

# SECTION 3

# PROCESS DESCRIPTION

A catalyst called Selectox is used for sulfur recovery from gas streams containing $H_2S$. This catalyst selectively oxidizes $H_2S$ to sulfur with air at low temperatures without forming $SO_3$ or oxidizing either hydrogen or light hydrocarbons. The process is entirely catalytic, eliminating the need for a thermal reaction furnace and resulting in a simplified Claus process. One-third of the $H_2S$ is oxidized catalytically with air to form $SO_2$, which then reacts with the remaining $H_2S$ to form elemental sulfur according to the principal Claus reactions.

$$H_2S + 3/2\ O_2 \longrightarrow SO_2 + H_2O$$

$$2H_2S + SO_2 \longrightarrow 3S + 2H_2O$$

The once-through process is called Selectox for acid gas feed streams containing less than 5% $H_2S$. For feed gas streams containing 5% $H_2S$ or greater, a recycle stream lean in $H_2S$ is used to limit the temperature rise from the exothermic reactions to a reasonable maximum considering the Claus chemical equilibrium and equipment metallurgy. This process is called Recycle Selectox.

The Selectox two-stage process illustrated in Figure 3-1 has a Selectox stage and one Claus stage whereas the Selectox three-stage process in Figure 3-2 has a Selectox stage and two Claus stages. The Recycle Selectox two-stage and three-stage processes in Figures 3-3 and 3-4 are identical to Figures 3-1 and 3-2, respectively, except for the addition of a blower to recycle a portion of Condenser No. 1 effluent back to Reheater No. 1.

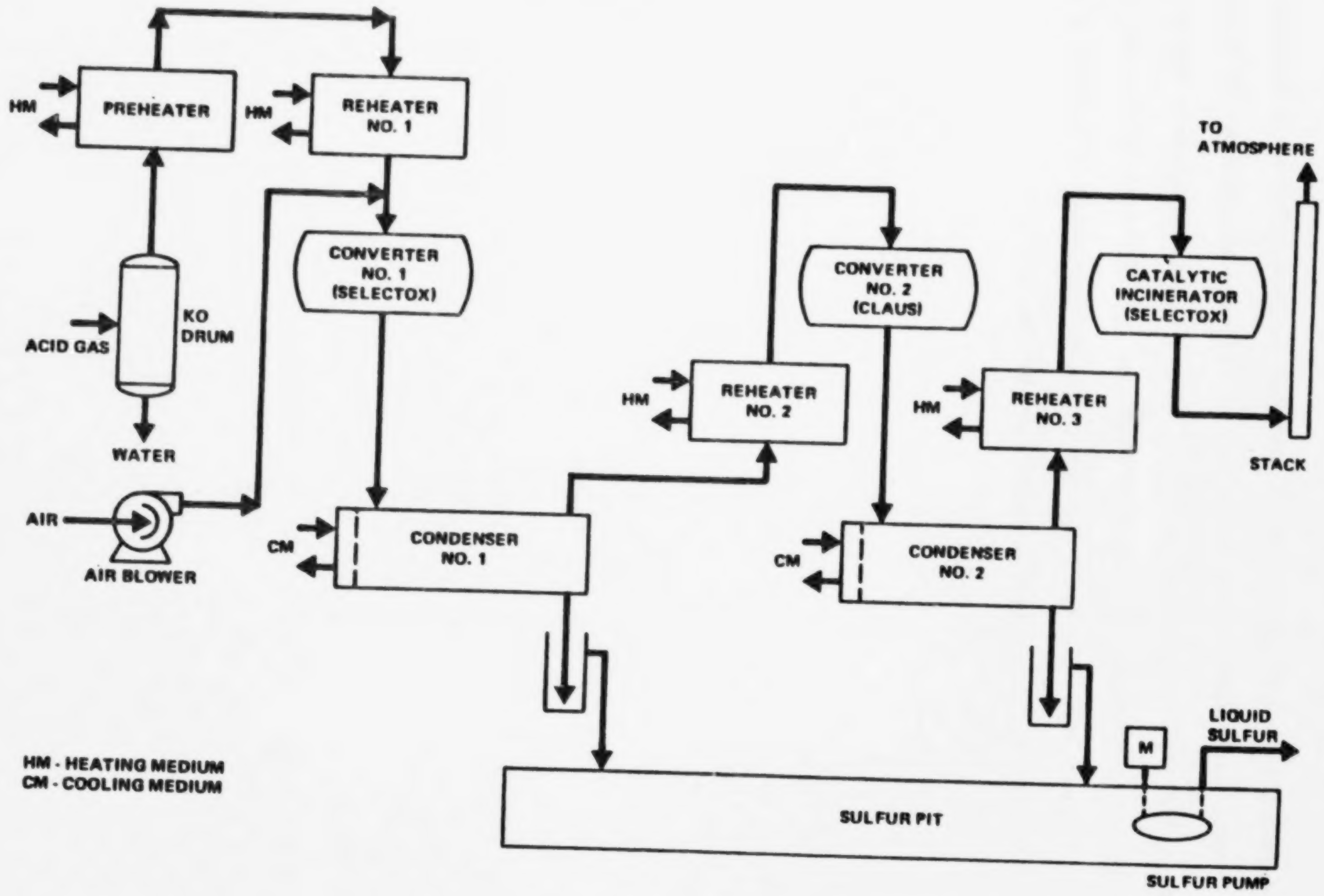
HM
PREHEATER
HM
REHEATER
NO. 1
ACID GAS
KO
DRUM
WATER
AIR
AIR BLOWER
CONVERTER
NO. 1
(SELECTOX)
CM
CONDENSER
NO. 1
HM
REHEATER
NO. 2
CONVERTER
NO. 2
(CLAUS)
CM
CONDENSER
NO. 2
HM
REHEATER
NO. 3
CATALYTIC
INCINERATOR
(SELECTOX)
TO
ATMOSPHERE
STACK
M
LIQUID
SULFUR
SULFUR PIT
SULFUR PUMP
HM - HEATING MEDIUM
CM - COOLING MEDIUM

Figure H-1. Selectox Two-Stage Process
(Figure 3-1)

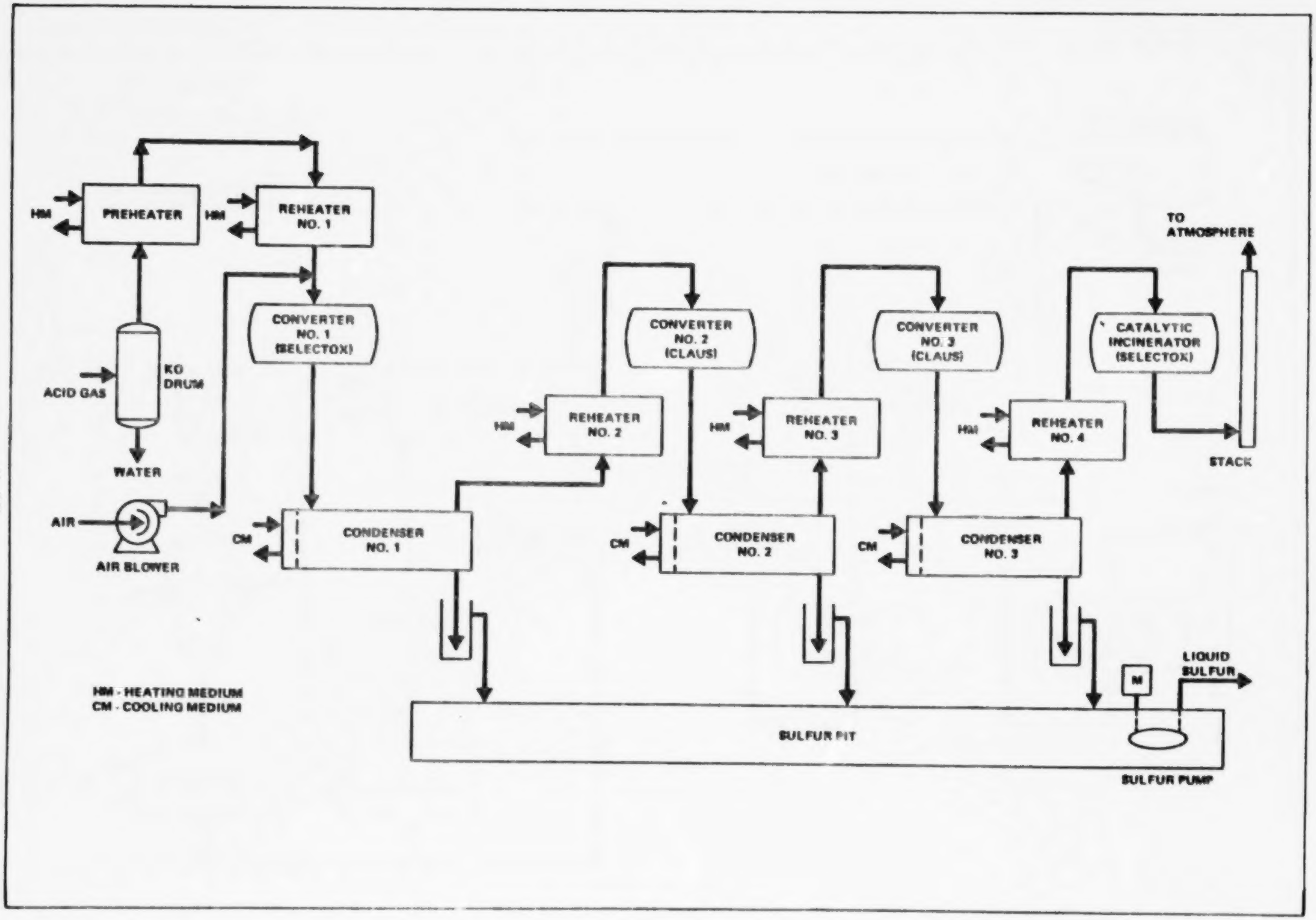
HM
PREHEATER
HM
REHEATER NO. 1
ACID GAS
KO DRUM
WATER
AIR
AIR BLOWER
CONVERTER NO. 1 (SELECTOX)
CM
CONDENSER NO. 1
HM
REHEATER NO. 2
CONVERTER NO. 2 (CLAUS)
CM
CONDENSER NO. 2
HM
REHEATER NO. 3
CONVERTER NO. 3 (CLAUS)
CM
CONDENSER NO. 3
HM
REHEATER NO. 4
CATALYTIC INCINERATOR (SELECTOX)
TO ATMOSPHERE
STACK
M
LIQUID SULFUR
SULFUR PIT
SULFUR PUMP
HM - HEATING MEDIUM
CM - COOLING MEDIUM

Figure H-2. Selectox Three-Stage Process
(Figure 3-2)

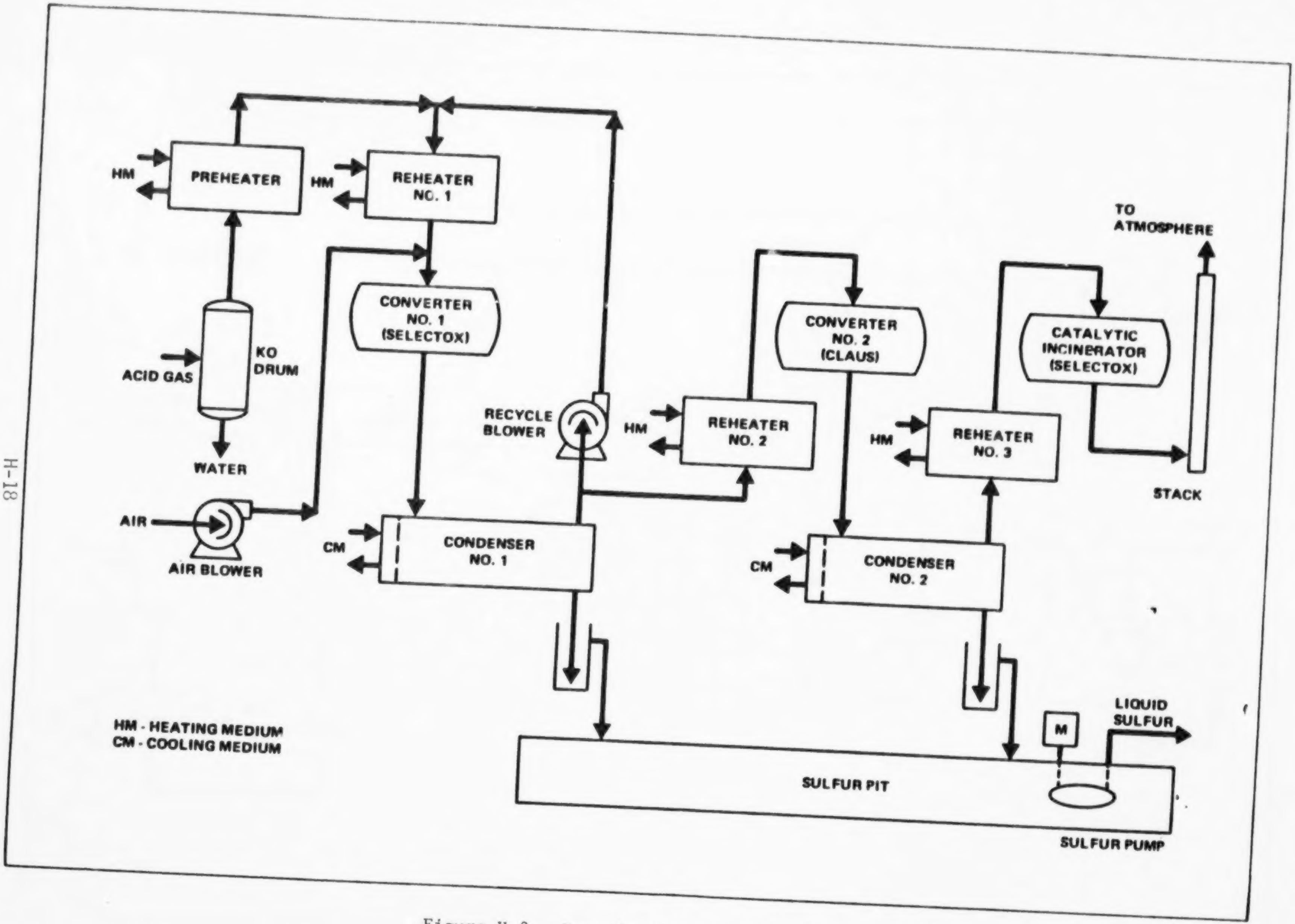
HM
PREHEATER
HM
REHEATER NO. 1
TO ATMOSPHERE
ACID GAS
KO DRUM
CONVERTER NO. 1 (SELECTOX)
CONVERTER NO. 2 (CLAUS)
CATALYTIC INCINERATOR (SELECTOX)
RECYCLE BLOWER
HM
REHEATER NO. 2
HM
REHEATER NO. 3
WATER
STACK
AIR
AIR BLOWER
CM
CONDENSER NO. 1
CM
CONDENSER NO. 2
HM - HEATING MEDIUM
CM - COOLING MEDIUM
M
LIQUID SULFUR
SULFUR PIT
SULFUR PUMP

Figure H-3. Recycle Selectox Two-Stage Process
(Figure 3-3)

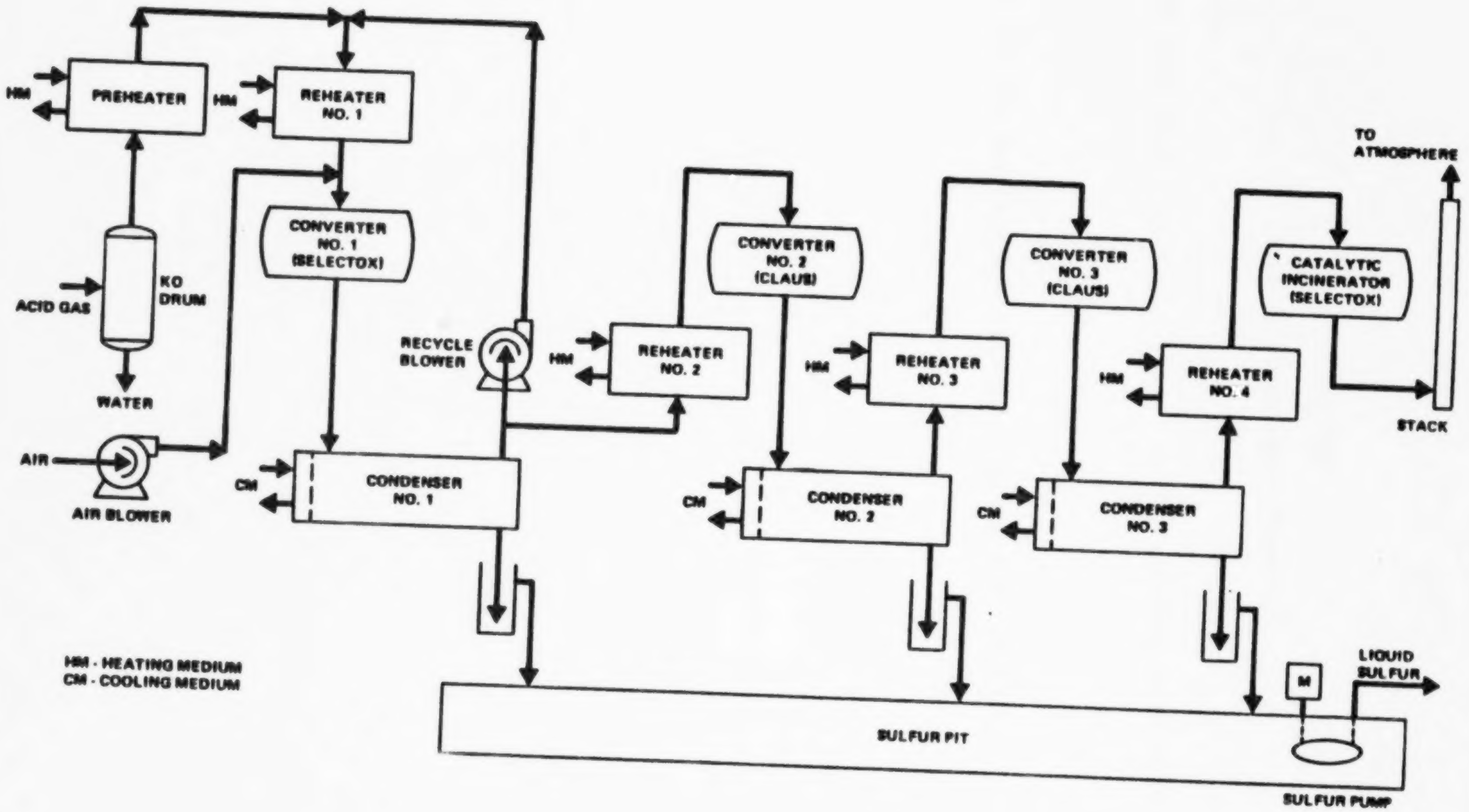
HM
PREHEATER
HM
REHEATER NO. 1
CONVERTER NO. 1 (SELECTOX)
ACID GAS
KO DRUM
WATER
AIR
AIR BLOWER
RECYCLE BLOWER
CM
CONDENSER NO. 1
HM
REHEATER NO. 2
CONVERTER NO. 2 (CLAUS)
CM
CONDENSER NO. 2
HM
REHEATER NO. 3
CONVERTER NO. 3 (CLAUS)
CM
CONDENSER NO. 3
HM
REHEATER NO. 4
CATALYTIC INCINERATOR (SELECTOX)
TO ATMOSPHERE
STACK
M
LIQUID SULFUR
SULFUR PIT
SULFUR PUMP
HM - HEATING MEDIUM
CM - COOLING MEDIUM

Figure H-4. Recycle Selextox Three-Stage Process
(Figure 3-4)

# SECTION 4

## DIRECT OPERATING COST DATA

The utilities requirements and the annual catalyst costs are provided in Table 4-1.

The sulfur recoveries as a percentage at the start-of-run and end-of-run and as average LT/D are shown in Table 4-2.

**Table** H-5 **- Utilities and Catalyst Costs**

(Table 4-1)

| Case No. | Sulfur Input (LT/D) | % $H_2S$ in Acid Gas | No. of Stages | Type | Electrical Power (kW) | Fuel Gas (MMBtu/hr) | Catalyst Cost ($/yr) |
|---|---|---|---|---|---|---|---|
| 1 | 0.5 | 2 | 2 | Selectox | 44 | 0.657 | 1,820 |
| 2 | 0.5 | 5 | 2 | Recycle Selectox | 46 | 0.277 | 880 |
| 3 | 0.5 | 12.5 | 2 | Recycle Selectox | 45 | 0.127 | 520 |
| 4 | 2.0 | 2 | 2 | Selectox | 60 | 2.630 | 7,280 |
| 5 | 2.0 | 5 | 2 | Recycle Selectox | 62 | 1.107 | 3,520 |
| 6 | 2.0 | 12.5 | 2 | Recycle Selectox | 60 | 0.507 | 2,070 |
| 7 | 2.0 | 2 | 3 | Selectox | 63 | 3.043 | 7,900 |
| 8 | 2.0 | 5 | 3 | Recycle Selectox | 62 | 1.306 | 3,770 |
| 9 | 2.0 | 12.5 | 3 | Recycle Selectox | 60 | 0.583 | 2,180 |
| 10 | 5.0 | 2 | 3 | Selectox | 99 | 7.609 | 19,730 |
| 11 | 5.0 | 5 | 3 | Recycle Selectox | 86 | 3.264 | 9,420 |
| 12 | 5.0 | 12.5 | 3 | Recycle Selectox | 93 | 1.456 | 5,450 |

**Table** H-6 - **Sulfur Recoveries**
(Table 4-2)

| Case No. | Sulfur Input (LT/D) | % $H_2S$ in Acid Gas | No. of Stages | Type | Start of Run (%) | End of Run (%) | Average (LT/D) |
|---|---|---|---|---|---|---|---|
| 1 | 0.5 | 2 | 2 | Selectox | 80.68 | 74.83 | 0.39 |
| 2 | 0.5 | 5 | 2 | Recycle Selectox | 87.90 | 81.65 | 0.42 |
| 3 | 0.5 | 12.5 | 2 | Recycle Selectox | 92.36 | 85.66 | 0.45 |
| 4 | 2.0 | 2 | 2 | Selectox | 80.68 | 74.83 | 1.56 |
| 5 | 2.0 | 5 | 2 | Recycle Selectox | 87.90 | 81.65 | 1.70 |
| 6 | 2.0 | 12.5 | 2 | Recycle Selectox | 92.36 | 85.66 | 1.78 |
| 7 | 2.0 | 2 | 3 | Selectox | 83.61 | 76.92 | 1.61 |
| 8 | 2.0 | 5 | 3 | Recycle Selectox | 91.40 | 84.00 | 1.75 |
| 9 | 2.0 | 12.5 | 3 | Recycle Selectox | 95.12 | 87.60 | 1.83 |
| 10 | 5.0 | 2 | 3 | Selectox | 83.61 | 76.92 | 4.01 |
| 11 | 5.0 | 5 | 3 | Recycle Selectox | 91.40 | 84.00 | 4.39 |
| 12 | 5.0 | 12.5 | 3 | Recycle Selectox | 95.12 | 87.60 | 4.57 |

OPERATING LABOR AND MAINTENANCE

For operating labor, assume 0.75 of an operator's time per shift and 0.25 of a supervisor's time per shift. The actual requirement will be somewhat less because these units need little operator attention.

Annual maintenance will be on the order of 3-1/2% of investment, including 1-1/2% for labor and 2% for investment.

Table H-7 presents annualized costs for 12 cases of sulfur feed rate/acid gas $H_2S$ concentration/Recycle Selectox process combinations.

Table H-7. ANNUALIZED COSTS FOR 12 CASES OF SULFUR FEED RATE/ACID GAS $H_2S$ CONCENTRATION/RECYCLE SELECTOX PROCESS COMBINATIONS[a]

| Case No. | Sulfur Feed Rate, LT/D | Mole Percent $H_2S$ in the Acid Gas, % | Sulfur Recovery Process | Cost $/year |
|---|---|---|---|---|
| 1 | 0.5 | 2.0 | Recycle Selectox 2-stage | 629,000 |
| 2 | 0.5 | 5.0 | Recycle Selectox 2-stage | 591,000 |
| 3 | 0.5 | 12.5 | Recycle Selectox 2-stage | 570,000 |
| 4 | 2.0 | 2.0 | Recycle Selectox 2-stage | 828,000 |
| 5 | 2.0 | 5.0 | Recycle Selectox 2-stage | 716,000 |
| 6 | 2.0 | 12.5 | Recycle Selectox 2-stage | 659,000 |
| 7 | 2.0 | 2.0 | Recycle Selectox 3-stage | 913,000 |
| 8 | 2.0 | 5.0 | Recycle Selectox 3-stage | 773,000 |
| 9 | 2.0 | 12.5 | Recycle Selectox 3-stage | 765,000 |
| 10 | 5.0 | 2.0 | Recycle Selectox 3-stage | 1134,000 |
| 11 | 5.0 | 5.0 | Recycle Selectox 3-stage | 788,000 |
| 12 | 5.0 | 12.5 | Recycle Selectox 3-stage | 637,000 |

---

[a] The costs calculated from the data presented in Sections 2 and 3 of this appendix, and Chapter 8, Table 8-3. Also refer to the docket (A-80-20-A), entry (II-B-44) for the step-by-step calculations.